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Physical-Chemical Treatment of A Municipal Wastewater Using Powdered Carbon. No. II

Envirotech Corp, Salt Lake City, Utah Eimco BSP Div

Prepared for

Municipal Environmental Research Lab, Cincinnati, Ohio Wastewater Research Div

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16. ABSTRACT

Salt Lake City municipal wastewater was treated in a nominal 100 gpm pilot plant by chemical coagulation-precipitation, powdered activated carbon adsorption and granular media filtration. Chemical-primary sludge was gravity thickened and vacuum filter dewatered. Spent carbon was gravity thickened, vacuum filter dewatered and thermally regenerated in a fluidized bed furnace. Solids-contact units were used for chemical treatment and carbon contacting. The pilot plant was operated over a 16-month period to demonstrate treatment effectiveness under diurnal flow conditions. Reuse of thermally regenerated carbon was practiced for 3 months.

Soluble organic materials were found to be removed by a combination of chemical coagulation, anaerobic biological activity and adsorption on powdered carbon. Spent powdered carbon was effectively regenerated with an average fixed carbon recovery of 90 percent. Lime and alum-primary sludges were effectively dewatered by vacuum filtration. A high quality effluent was consistently produced, similar to that expected for conventional biological treatment followed by tertiary treatment for phosphorus and suspended solids removal.

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nitrogenous substances as pollutants or health hazardous compounds. Therefore it was attempted to compare nitrate reducing microorganisms in their reactions to different environmental conditions in laboratory experiments. Changing oxygen concentration, pH, temperature, nitrate or nitrite concentration affected differently the denitrification process of various isolated microorganisms. Biological and chemical factors were evaluated during nitrite transformation in autoclaved and non-autoclaved soil by determination of the evolvement of nitrogenous gases. During chemical nitrite volatilization, which occurred essentially at a low pH, the major gases evolved were nitric oxide and nitrogen dioxide, but if biological activity was predominant in a neutral and alkaline environment, nitrous oxide and molecular nitrogen were formed. The validity of laboratory observations in relation to field studies in the domain of denitrification is discussed and evaluated. (Modified author abstract)

PB-224 471/3GA PC$3.00/MF$1.45
Spartanburg County Planning and Development Commission, S.C.
INITIAL RECONNAISSANCE SURVEY FOR THE TOWN OF LANDRUM, SPARTANBURG COUNTY, SOUTH CAROLINA.
May 73, 12p SC-40-0014-0743

Descriptors: (*Urban planning, *South Carolina), Population growth, Recommendations, Central city, Project planning, Residential buildings, Urban transportation, Finance.
Identifiers: *Landrum(South Carolina).

The report describes a standard reconnaissance survey conducted in accordance with HUD 701 requirements and designed to identify the problems and needs of Landrum, SC.

PB-224 472/1GA PC$8.00/MF$1.45
New York State Office of Planning Services, Albany.
PLANNING AND REDEVELOPMENT PROGRAM, PART I. VILLAGE OF PAINTED POST, NEW YORK.
Final rept.
Jun 73, 20p

Descriptors: (*Regional planning, *New York), Community development, Urban renewal, Building codes, Law enforcement, Residential buildings, Financing, Relocation, Floods.
Identifiers: *Painted Post(New York).

The report is the first of two parts of the village's planning and redevelopment program. The report contains material prepared for the village of Painted Post, New York in order to meet federal urban renewal certification requirements. The information presented is a background analysis of existing conditions and problems and is used as a basis for the workable program. It includes the following specific studies: Codes and ordinances; neighborhood analysis; administrative organization; financing; and housing for displaced families. (Modified author abstract)

PB-224 473/9GA PC$6.25/MF$1.45
Waccamaw Regional Planning and Development, Council, Georgetown, S.C.
AN ENVIRONMENTAL, HISTORICAL AND RECREATIONAL ATLAS OF THE WACCAMAW REGION.
Sep 73, 84p SC-04-0014-0653

Descriptors: (*Regional planning, *South Carolina), Resources, Site surveys, Recreational facilities, Environmental surveys, History, Maps.
Identifiers: Georgetown County(South Carolina), Horry County(South Carolina), Williamsburg County(South Carolina).

The report provides a general description of historical, recreational and environmental sites within the Waccamaw Region, SC and locates each of these sites on a map. (Author)

PB-224 476/2GA PC-GPO/MF$1.45-NTIS
Kansas State Univ., Manhattan. Dept. of Civil Engineering.
PILOT PLANT DEMONSTRATION OF A LIME-BIOLOGICAL TREATMENT PHOSPHORUS REMOVAL METHOD.
Lawrence Schmid. Jun 73, 59p W74-00155, EPA-R2-73-159
Paper copy available from GPO $0.85 as EP1.23/2:73-159 and EPA-R2-73-159.

Descriptors: (*Chemical removal(Sewage treatment), *Phosphorus), Pilot plants, Sewage treatment, Activated sludge process, Performance evaluation, Process charting, Calcium oxides, Precipitation(Chemistry), Coagulation, PH, Biochemical oxygen demand.
Identifiers: Water pollution control.

A 15,000 gpd pilot plant was constructed to demonstrate the capabilities of a lime treatment process for phosphorus removal. The lime treatment of raw wastewater removes the bulk of the phosphorus, and a subsequent biological process removes an additional increment of phosphorus via cell synthesis. The pilot plant used for the study was a package-type, prefabricated unit. Additional small tanks were provided for sludge storage and measurement. A variable speed motor was mated to a program controller to duplicate a diurnal flow pattern through the pilot plant. Lime was introduced into the raw wastewater in response to an automated pH control system. Instrumented systems were also developed for turbidity and ortho-phosphate concentrations in the plant effluent. The raw wastewater entering the pilot plant was obtained from the wet well of the Manhattan, Kansas Municipal Plant. (Modified author abstract)

PB-224 477/0GA PC$6.25/MF$1.45
Massachusetts Inst. of Tech., Cambridge. Ralph M. Parsons Lab. for Water Resources and Hydrodynamics.
METHODOLOGY FOR ASSESSING THE POTENTIAL IMPACT OF URBAN DEVELOPMENT ON URBAN RUNOFF AND THE RELATIVE EFFICIENCY OF RUNOFF CONTROL ALTERNATIVES,
Guy Leclerc, and John C. Schaake, Jr. Mar 73, 256p 167, R73-23 W74-00001, OWRR-C-2137(3403)(5)
Contract DI-14-31-0001-3403

Descriptors: (*Surface water runoff, *Urbanization), Urban areas, Methodology, Estimating, Stochastic processes, Rainfall, Computerized simulation, Watersheds, Fluid infiltration, Frequency distribution, Water storage.
Identifiers: *Urban hydrology, Storm water runoff.

A methodology was developed to assess the potential impact of urban development on urban runoff and to measure the relative efficiency of runoff control alternatives. The methodology utilizes runoff frequency curves, derived at different stages of development of the urban catchment. Comparisons of these curves completely quantify the impact of urban development and/or the efficiency of a control structure. Runoff frequency curves are derived directly from the rainfall process. The method of solution utilizes a stochastic model of the rainfall process and a deterministic model of the catchment. The method of solution, referred to as stochastic/deterministic simulation, incorporates the stochastic rainfall model, the deterministic catchment model, and results of the infiltration estimator. Results show that the solution procedure and the methodology developed are practical and are recommended for urban drainage analysis. (Modified author abstract)

PB-224 494/5GA PC-GPO/MF$1.45-NTIS
Envirotech Corp., Salt Lake City, Utah. Eimco Process Machinery Div.
PHYSICAL-CHEMICAL TREATMENT OF A MUNICIPAL WASTEWATER USING POWDERED CARBON.
Environmental protection technology series.
Donald E. Burns, and Gerald L. Shell. Aug 73, 237p W74-00154, EPA-R2-73-264
Contract EPA-14-12-585
Paper copy available from GPO $2.25 as stock no. EP1.23/2:73-264.

Descriptors: (*Chemical removal(Sewage treatment), *Sewage treatment), (*Activated carbon treatment, Sewage treatment), Pilot plants, Coagulation, Precipitation(Chemistry), Dewatering, Alums, Calcium oxides, Iron chlorides, Sludge disposal, Cost estimates, Polyelectrolytes, Vacuum filtration, Regeneration(Engineering).
Identifiers: *Physical chemical treatment, Water pollution control, Salt Lake City(Utah).

A municipal wastewater was treated in a nominal 100 gpm pilot plant by chemical coagulation-precipitation, powdered activated carbon adsorption and granular media filtration. Spent carbon was gravity thickened, vacuum filter dewatered and thermally regenerated in a fluidized bed furnace. Solids-contact units were used for chemical treatment and carbon contacting. Ferric chloride, alum or lime were all found to effectively produce coagulation and phosphorus insolubilization. Based on total treatment costs, including sludge disposal, alum treatment was estimated to be the economic choice for Salt Lake City municipal wastewater. Organic removal in the powdered carbon contactors was substantially enhanced by anaerobic biological activity. The use of solids-contact treatment units for carbon contacting resulted in effecting gravity clarification without the use of chemicals. The powdered carbon physical-chemical treatment system produced a treated effluent similar to that expected for biological treatment followed by tertiary treatment for phosphorus removal. Carbon losses of 17 to 60 percent were experienced across the fluidized bed furnace regeneration system. The cause of high carbon losses was identified as ignition of carbon instead of gas which was injected into the fluidized bed to scavenge excess oxygen. (Author)

PB-224 500/9GA PC$8.00/MF$1.45
Cornell Univ., Ithaca, N.Y. Water Resources and Marine Sciences Center.
STUDIES IN THE ANALYSIS OF METROPOLITAN WATER RESOURCE SYSTEMS. VOLUME VI. ESTIMATING ECONOMIES OF SCALE IN THERMAL ELECTRIC POWER SYSTEMS SUBJECTED TO ENVIRONMENTAL QUALITY CONSTRAINTS.
Technical rept.,
Charles R. Scherer. Aug 73, 395p TR-67 W74-00002, OWRR-C-3101(3684)(2)
Contract DI-14-31-0001-3684
See also Volume 5 dated Oct 72, PB-217 862.

Descriptors: (*Electric power generation, *Economic analysis), (*Water pollution, Electric power generation), Steam electric power generation, Cost estimates, Operating costs, Transmission loss, Electric power transmission, Air pollution, Benefit cost analysis, Electric power plants, Capitalized costs, Standards, Optimization, Mathematical models, Water pollution.
Identifiers: Economies of scale.

This study is concerned with the estimation of economies of scale in thermal electric power systems subjected to water quality standards. The approach is engineering-economic. Mixed-integer programming is used to solve a mathematical model of a power system subject to constraints defining (1) the load duration curves of several demand centers located on a spatial grid or network, (2) endogenously determined reserves equal to the largest two generating units in the system, (3) nonlinear transmission losses, and (4) environmental

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PHYSICAL-CHEMICAL TREATMENT OF A MUNICIPAL WASTEWATER USING POWDERED CARBON

No. II

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Contract No. 68-01-0183

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MUNICIPAL ENVIRONMENTAL RESEARCH LABORATORY
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DISCLAIMER

This report has been reviewed by the Municipal Environmental Research Laboratory--Cincinnati, Ohio, U. S. Environmental Protection Agency, and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the U. S. Environmental Protection Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

This report summarizes the results of a pilot plant program which developed the use of chemical clarification and powdered activated carbon adsorption for the treatment of municipal wastewater. The development of such alternative processes provides additional tools for use by water pollution control agencies in their efforts to maintain and improve the quality of the environment.

Francis T. Mayo
Director
Municipal Environmental
Research Laboratory
Cincinnati

ABSTRACT

Salt Lake City municipal wastewater was treated in a nominal 100 gpm pilot plant by chemical coagulation-precipitation, powdered activated carbon adsorption and granular media filtration. Chemical-primary sludge was gravity thickened and vacuum filter dewatered. Spent carbon was gravity thickened, vacuum filter dewatered and thermally regenerated in a fluidized bed furnace. Solids-contact units were used for chemical treatment and carbon contacting.

The pilot plant was operated over a 16 month period to demonstrate treatment effectiveness under diurnal flow conditions. Reuse of thermally regenerated carbon was practiced for 3 months.

A high quality effluent was consistently produced, similar to that expected for conventional biological treatment followed by tertiary treatment for phosphorus and suspended solids removal.

Spent powdered carbon was effectively regenerated with fixed carbon recovery of 90 percent on the average.

Soluble organic materials were found to be removed by a combination of chemical coagulation, anaerobic biological activity and adsorption on powdered carbon.

Lime and alum-primary sludges were effectively dewatered by vacuum filtration.

This report was submitted in fulfillment of Project No. 17020 HNH, Contract No. 68-01-0183 by the Eimco BSP Division of Envirotech Corporation under the sponsorship of the Environmental Protection Agency. Pilot plant work was completed December, 1973.

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The support, counsel and patience of the project officer, Mr. James Westrick and Mr. Jesse M. Cohen of the Wastewater Research Division, Municipal Environmental Research Laboratory, Cincinnati, Ohio, Environmental Protection Agency, is acknowledged with sincere thanks.

SECTION I

CONCLUSIONS

With the completion of this study essentially all of the originally defined project objectives were achieved. Specific pertinent conclusions deduced from the results of this study are as follows:

1. When employing alum or $FeCl_3$ pretreatment, two stage counter-current powdered carbon contacting (at 100 mg/ℓ carbon) and typical diurnal flow conditions, the powdered activated carbon - physical chemical treatment (PAC-PCT) process evaluated consistently produced a high quality effluent of less than 5 mg/ℓ chemical oxygen demand, 5 mg/ℓ suspended solids and 0.3 mg/ℓ phosphorus.

2. When properly operated, the solids-contact treatment units employed in the chemical and powdered carbon contacting-clarification steps performed very well during diurnal flow conditions.

3. Suitable design and operating conditions were identified for the critical final liquid-solids separation step of granular media filtration as evidenced by long term production of highly clarified pilot plant effluent.

4. Powdered carbon regeneration using a fluidized bed furnace (FBF), with off-gas recycle, resulted in fixed carbon recoveries, on the average, in excess of 90 percent.

5. Thermally regenerated powdered carbon did not exhibit full recovery of adsorptive capacity but when reused in the PAC-PCT process no significant loss of treatment effectiveness was found.

6. Acceptable recoveries of fixed carbon was not achieved with the FBF unit used when employing the originally designed bed injection gas principal of operation.

7. Soluble organics (COD) were found to be removed by three mechanisms in the PAC-PCT process; namely, chemical coagulation, anaerobic biological activity and adsorption on activated carbon.

8. For the system studied, use of alum or ferric chloride in the chemical treatment step resulted in significant removal of soluble COD, whereas use of lime did not.

9. No significant treatment or cost benefit of counter-current staging of powdered carbon contacting was precisely defined.

10. Lime and alum-primary sludge mixtures, produced in the chemical treatment step, were found to be effectively concentrated by gravity thickening and dewatered by vacuum filtration.

11. Ferric-primary sludge thickened well but was found very difficult to dewater by vacuum filtration.

12. Laboratory and pilot plant gravity thickening test results were not consistently correlated. Short comings in the laboratory procedures and/or inconsistent operation of the pilot plant units were considered primary causes.

13. Laboratory leaf tests and pilot plant vacuum filtration test results were found to be precisely correlated.

14. The use of ultraviolet light absorption monitoring of carbon system feed or effluent was not found to provide useful real-time operational intelligence.

15. Alum-powdered carbon treatment of 10 mgd of Salt Lake City municipal wastewater, with thermal regeneration and reuse of carbon, to produce an effluent containing less than 5 mg/ℓ COD, 5 mg/ℓ SS and 0.5 mg/ℓ phosphorus was estimated to cost 33.4 ¢/1000 gal. If powdered carbon was not regenerated but used only once, the total treatment cost was estimated to be 36.2 ¢/1000 gal.

SECTION II

RECOMMENDATIONS

1. It is strongly recommended that the Environmental Protection Agency (EPA) promote, and if necessary, fund construction and preliminary operation of a 3 to 10 mgd PAC-PCT plant to treat a municipal wastewater similar to Salt Lake City, Utah. Such a project would commercialize PAC-PCT technology developed by the EPA and others by providing full scale demonstration of treatment effectiveness and also allow determination of any undefined operational or performance problems.

   In addition, if a flexible thermal regeneration system were constructed, full scale optimization of powdered carbon regeneration could be accomplished.

2. It is recommended that a cost sensitivity study of the PAC-PCT process be conducted to identify cost sensitive areas which might be justifiable future development efforts.

3. It is recommended that chemical-primary sludge dewatering alternatives be evaluated on a real time basis. Such alternatives should include, at least, vacuum filtration, centrifugation and pressure filtration.

4. It is recommended that a pilot plant study be conducted comparing parallel granular and powdered carbon systems following chemical treatment-clarification of a raw municipal wastewater.

SECTION III

INTRODUCTION

Physical-Chemical Treatment (PCT) of wastewater, as discussed herein, consists of chemical coagulation-precipitation followed by powdered activated carbon (PAC) contacting. PCT was originally developed as a tertiary method for reclamation of secondary biological effluents. In recent years, PCT has been evaluated as a method for treatment of raw wastewaters. Because of well developed regeneration technology, major emphasis has been placed on studying granular activated carbon (GAC) contacting. The use of PAC has several real and potential advantages wh[illegible] compared to GAC.[1] They are as follows:

1) The cost of powdered carbon is substantially less.

2) Powdered carbon adsorption will equilibrate with soluble wastewater organics in a small fraction of the time.

3) Powdered carbon might be supplied on demand to meet varying feed organic strength.

4) Powdered carbon systems require a fraction of the carbon inventory.

5) Powdered carbon contacting systems are more amenable to control of undesirable biological activity.

Environmental Protection Agency (EPA) funded studies have indicated that thermal regeneration of PAC is technically feasible and may be economically justified.[2,3] In 1969, Eimco Division of Envirotech Corporation and EPA negotiated a contract to conduct an extensive pilot plant study for the purpose of demonstrating the operability and treatment effectiveness of the PAC-PCT process applied to a municipal wastewater. A nominal 100 gpm constant flow pilot plant was constructed in Salt Lake City, Utah, in

1969. This pilot plant was operated during 1970 and the first half of 1971. The broad objectives of this study were as follows:

1) To determine the relative treatment and cost effectiveness of lime, alum and ferric chloride chemicals for precipitating phosphorus and removal of suspended solids in solids-contact treatment units.

2) To determine the effect of carbon dosages and number of counter-current stages of contacting in solids contact treatment units on the removal of soluble chemical oxygen demand (SCOD).

3) To demonstrate the effectiveness of carbon regeneration by a fluidized bed furnace system and the reuse effectiveness of regenerated carbon.

4) To compile process operating and efficiency data suitable for design and economic evaluation of a full scale PAC-PCT plant.

Most of the above objectives were achieved during this study. Lime, alum and ferric chloride were all found to be effective in the chemical pretreatment step. Alum treatment of 10 mgd of Salt Lake City wastewater for phosphorus removal and clarification was estimated to be less expensive than ferric chloride or lime treatment. The PAC system effluent SCOD was found to be relatively insensitive to the carbon dosage employed. Use of two-stage counter-current contacting was found to require less carbon than single-stage contacting. However, precision of pilot plant test results was such that a high degree of statistical significance was not determined for this difference. Regeneration of spent carbon in the fluidized bed furnace system resulted in about 40 to 60 percent loss of carbon.

The results of this study demonstrated the operability and treatment effectiveness of the PAC-PCT process. Estimated economics indicated that the process may be an economic alternative to presently available treatment approaches for certain wastewaters. However, several areas needed further study to develop new and/or more reliable design data. A follow-on contract was negotiated for an additional year of studies based on those areas needing further work. Due to the unsuccessful performance of the fluidized bed carbon regeneration system, as originally designed and installed, a 3-month contract extension was negotiated. The fluidized bed furnace was slightly modified, allowing achievement of efficient regeneration and the opportunity to recycle and re-use regenerated carbon.

OBJECTIVES

Objectives of the current study were as follows:

1) Demonstration of treatment effectiveness by long term pilot plant operation under diurnal flow conditions. Specifically, the effect of diurnal flow on the performance of the solids-contact treatment units was to be determined. It was also desired to demonstrate the long term consistency of pilot plant effluent quality.

2) Continued development of the fluidized bed furnace system to obtain fixed carbon losses of less than 25 percent. The quality of regenerated carbon, as determined by re-use in the pilot plant, was to be determined.

3) Obtain pilot plant performance results for gravity thickening and vacuum filtration dewatering of chemical-primary sludges (lime, alum and ferric chloride). Laboratory thickening and vacuum filter leaf test results would also be obtained. Correlations between laboratory and pilot plant results would be defined.

4) Evaluate the utility of certain continuously monitoring water quality evaluation instruments for assisting operation of the PAC-PCT process. Specific instruments to be used would be ultra violet light absorption, orthophosphate, turbidity and pH measuring devices.

APPROACH

The contract work was divided into two phases of operation. The first phase was to last approximately 3 months. During this time, the pilot plant, which had been dormant for approximately 9 months, was renovated and modified. Modification of existing facilities and some new construction were completed. In addition, process evaluation and control instruments were purchased and installed.

After the pilot plant was made operable, several weeks were devoted to training plant operating personnel. During the remainder of Phase I, the pilot plant was operated under constant flow conditions with lime pretreatment and single-stage carbon contacting. Major emphasis was placed on producing spent carbon for regeneration operations. The purpose was to expedite development work on the fluidized bed furnace system.

Phase II of the contract work comprised a 12-month period which included a 3-month extension. The major purpose of Phase II plant operation was to obtain long term operational and performance data. The pilot plant was operated under diurnal flow conditions with various peak to minimum flow ratios. During three major pilot plant operating campaigns, which were approximately 3 months each, either lime, alum or ferric chloride pretreatment of the raw municipal wastewater was used. This schedule was to allow near equal compilation of continuous gravity thickening and intermittent vacuum filtration data on the chemical-primary sludges.

Two-stage counter-current contacting of the carbon was used during most of Phase II, since it had previously been identified as the most economical mode of operation.[1] A carbon dosage of about 100 mg/ℓ was used. If pilot plant effluent SCOD concentrations of less than 15 to 20 mg/ℓ had not been obtained, then a higher carbon dosage of approximately 300 mg/ℓ would have been used.

Granular media filter bed design would be evaluated in an attempt to maximize efficiency of filter operation.

A major activity during Phase II would be the compilation of carbon regeneration and re-use data. Of course, this activity had to be preceeded by successful development of the fluidized bed furnace system.

SECTION IV

PILOT PLANT FACILITIES

A photo of the PAC-PCT pilot plant facility is shown on Figure 4-1. The liquid treatment process flow diagram for this facility is shown on Figure 4-2.

Raw Salt Lake City municipal wastewater was chemically treated to precipitate phosphorus and coagulate suspended solids. The chemically treated wastewater was solids-contacted and gravity settled. The clarified wastewater constituted the PAC treatment system feed.

Chemical contactor underflow sludge was sent to a chemical-primary sludge treatment system. This system consisted of continuous gravity thickening, thickened sludge storage and intermittent vacuum filtration.

Carbon was used for the sole purpose of removing soluble organic material. The carbon contacting and clarification system consisted of single or two-stage counter-current contacting in solids-contact treatment units followed by granular media filtration. Spent carbon was removed as carbon contactor underflow and discharged to a gravity thickener. Thickened spent carbon was stored prior to being thermally regenerated in a fluidized bed furnace system. All carbon containing streams were recycled to the first-stage carbon contactor (e.g., thickener overflow, vacuum filter filtrate and granular media filter backwash water).

The following subsections are a detailed description of the major components of the pilot plant.

RAW WASTEWATER FLOW

The pilot plant influent was coarse screened and comminuted Salt Lake City raw wastewater. About 350 gpm was diverted from the Salt Lake City pump station discharge header to the pilot plant raw wastewater sump. Flow not pumped to the pilot plant was returned to the pump station wet well through a gravity overflow. During the first 7 months of pilot plant

FIGURE 4-1: OVERALL PHOTO OF PAC-PCT PILOT PLANT

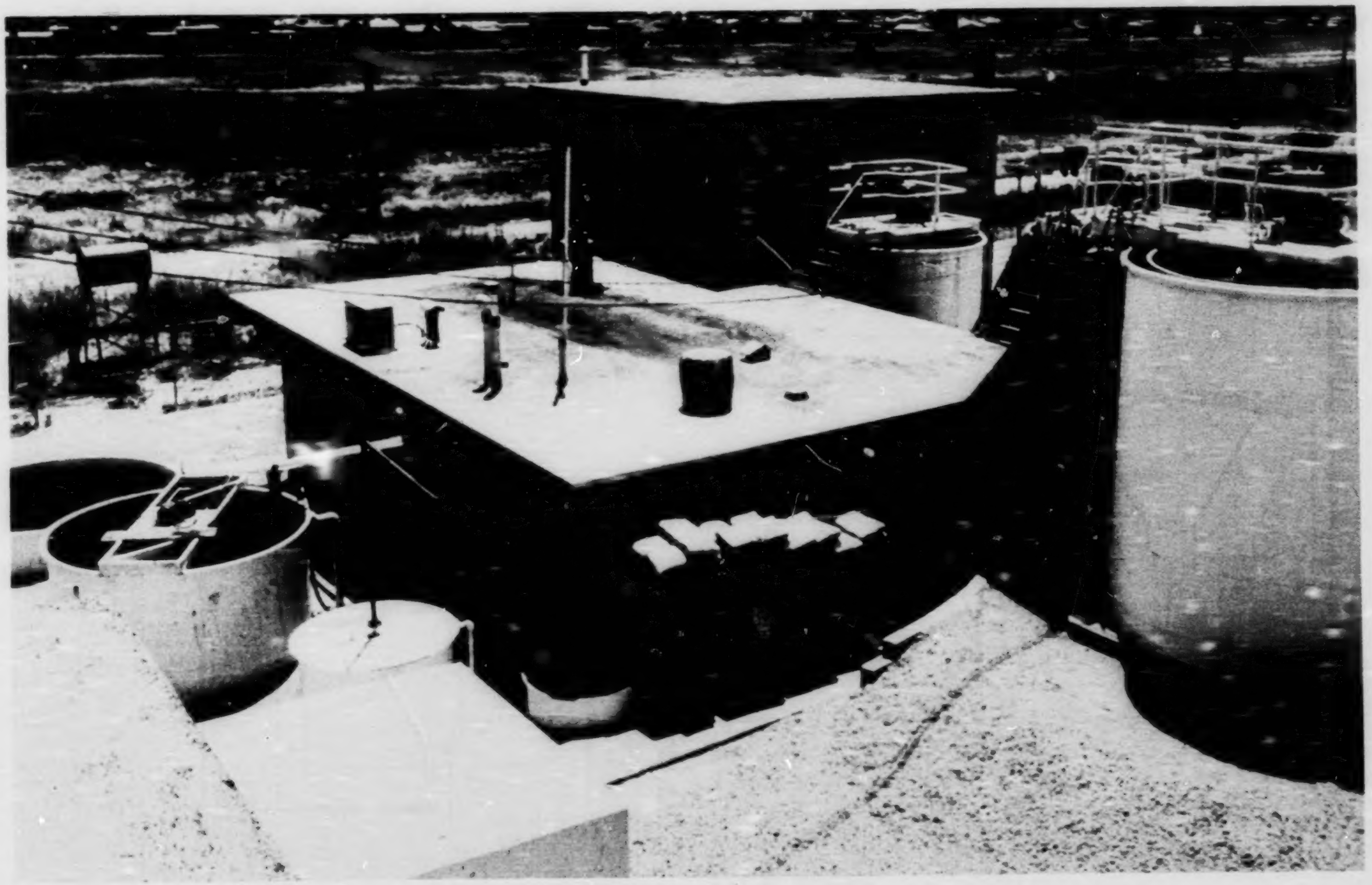
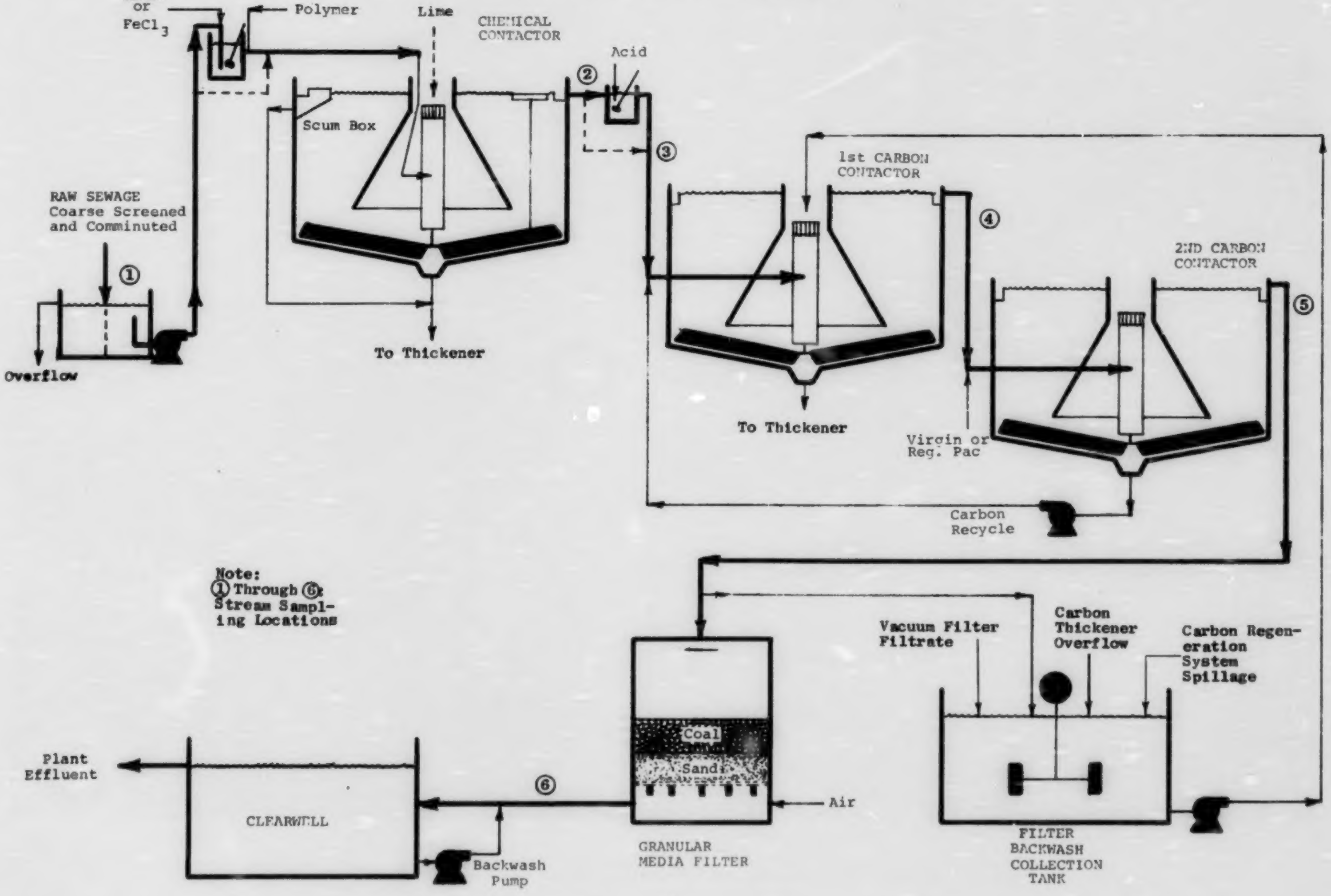


FIGURE 4-2:
PROCESS FLOW DIAGRAM


Alum or $FeCl_3$
Polymer
Lime
CHEMICAL CONTACTOR
Scum Box
RAW SEWAGE Coarse Screened and Comminuted
①
Overflow
To Thickener
②
Acid
③
1st CARBON CONTACTOR
④
To Thickener
2ND CARBON CONTACTOR
⑤
Virgin or Reg. Pac
Carbon Recycle
Note: ① Through ⑥: Stream Sampling Locations
Vacuum Filter Filtrate
Carbon Thickener Overflow
Carbon Regeneration System Spillage
Coal
Sand
Air
GRANULAR MEDIA FILTER
FILTER BACKWASH COLLECTION TANK
Plant Effluent
CLEARWELL
⑥
Backwash Pump

operation, all rags and most of the grit in the raw wastewater were allowed to go to the chemical treatment unit. Because of rag clogging problems in the thickened sludge handling system pumps, a "de-ragging" screen was placed in the raw wastewater sump for the last 8 months of operation. This screen had 1/2-inch square openings.

The raw wastewater pump was sized for a flow in excess of 200 gpm. It was driven by a 7-1/2 Hp, 180-volt, DC shunt wound motor. The motor received a 0-180 volt signal from an electronic control device, which inturn received a signal from a cam-operated pneumatic-electrical controller. The cam was on a 24-hour clock. The cam shape was a variable allowing any pattern of diurnal flow variation desired. The actual pattern of flow used was identical to that for Salt Lake City municipal wastewater.

Flow from the raw wastewater pump to the chemical treatment unit passed through a magnetic flow meter. A signal from this flow sensor was transmitted to a flow indicating, recording, totalizing and control unit.

CHEMICAL TREATMENT SYSTEM

Chemical Treatment Unit

Chemical treatment and clarification of raw wastewater was accomplished in a 12-ft diameter by 10.5-ft side water depth Reactor-Clarifier treatment unit.*

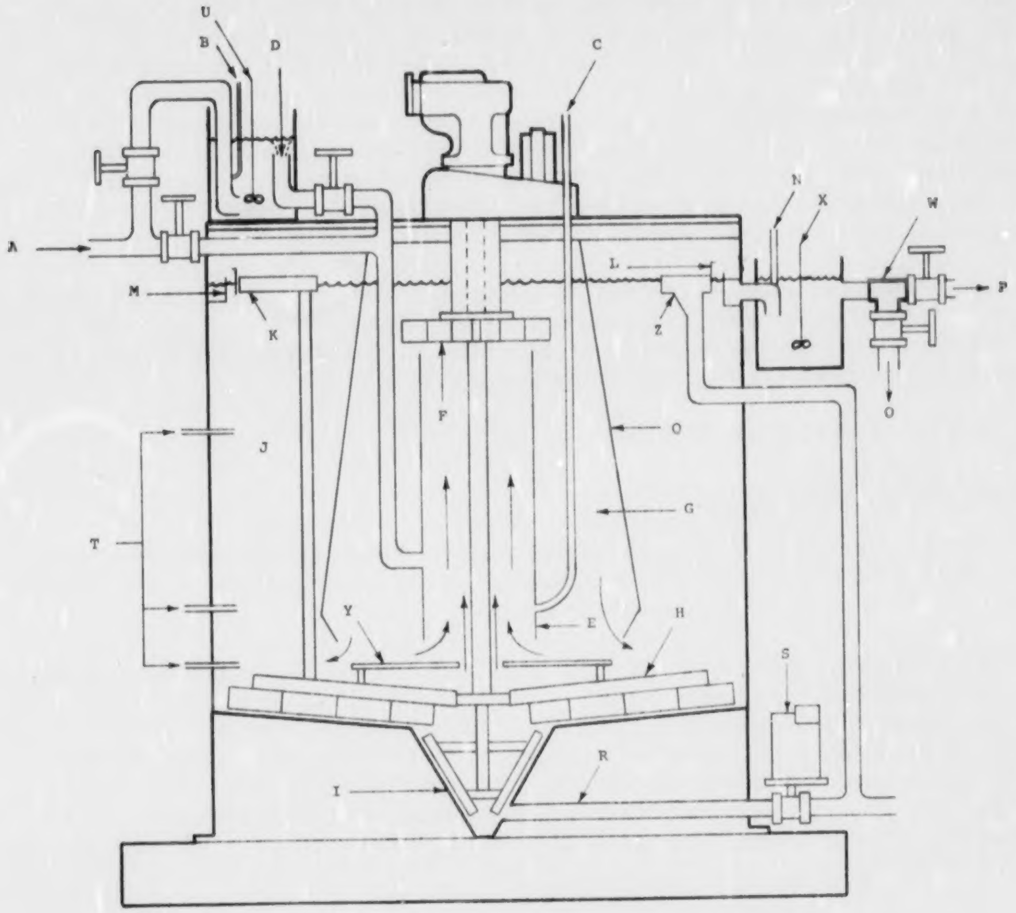
Alum or $FeCl_3$ was fed to the wastewater at B and flash mixed at U (capital letters refer to locations on Figure 4-3). If polyelectrolyte was used it was added at D. When lime treatment was used, the raw wastewater flow bypassed the flash-mix tank and went directly to the draft tube E. Lime was added at C. This resulted in the lime contacting copious amounts of recycled solids simultaneously with the raw wastewater.

A slow speed turbine F pumped raw wastewater, chemicals and previously formed chemical precipitates up the draft tube and out into the reaction zone. The pumping turbine capacity was adjustable in the range of about 5 to 12 times the average hydraulic throughput of the unit. Reaction zone slurry

---

*Manufactured by Eimco BSP, Division of Envirotech Corporation, Salt Lake City, Utah

FIGURE 4-3:
EIMCO REACTOR-CLARIFIER, SOLIDS-CONTACT CHEMICAL TREATMENT UNIT



SYMBOLS

A. Influent Pipe
B. Inorganic Coagulant Feed
C. Lime Feed
D. Polymer Feed
E. Draft Tube
F. Pumping Turbine
G. Sample Point, Reaction Zone
H. Sludge Rake
I. Sludge Thickening Cone
J. Expanding Area Clarification Zone
K. Liquid Surface Skimmer
L. Float Baffle
M. Peripherial V-Notched Launder
N. Acid Feed Point
O. Effluent to First Carbon Contactor
P. Effluent to Second Carbon Contactor or Drain
Q. Reaction Zone Cone
R. Sludge Blowdown Line
S. Timer Controlled Sludge Valve
T. Sample Taps
U. Chemical Addition Flash Mixer
W. Flow Splitter Box, Effluent Sample Point
X. Acid Neutralization Flash Mixer
Y. Baffle Plate
Z. Scum Box

flowed under the reaction zone cone Q into an increasing area upflow clarification zone. Settled solids were moved to the center of the unit by sludge rake H and either drawn up the draft tube or removed through the thickening cone I. Partially thickened sludge was automatically removed from the treatment unit by gravity flow through line R, and an air operated, timer controlled sludge valve S. Chemical sludge blowdown was routed to the chemical sludge handling system.

Since the chemical treatment unit received raw wastewater, provision was made for clarification zone surface skimming by mechanism K. Floatable materials were kept from passing over the peripherial launder M (a V-notched weir) by baffle L. Effluent neutralization was accomplished by addition of concentrated sulfuric acid to the effluent in a flash mix tank at location X. Clarified effluent exited the unit at splitter box W, which divided flow to one or two carbon contactors or to drain. A sample of clarifier effluent was continuously withdrawn at point X and conveyed by gravity flow to a composite sampler and process turbidimeter.

Except for the following modifications, the solids-contact unit was unchanged from that used in previous contract work. The original square turbine blades were replaced with long rectangular blades which extended from the periphery of the draft tube to next to the rake drive shaft. This modification eliminated ragging problems previously experienced.

During alum and ferric chloride treatment campaigns, a baffle plate Y was installed in the solids-contact units. It was anticipated that this baffle would hydraulically isolate the turbulent reaction zone and sluge outlet region. Higher underflow sludge concentration than was experienced during previous contract work was expected.

Flash mixing units were installed to improve the reaction of inorganic coagulants and neutralizing acid with the wastewater. During previous contract work coagulant was injected into the influent pipe just before the draft tube and acid was added to the launder 180 degrees from the effluent splitter box.

The sludge blowdown line R was changed from a 2 to a 4-inch diameter pipe. This modification was for the purpose of minimizing clogging of the smaller line with rags, grit and foreign objects.

Chemical Sludge Handling System

The chemical sludge handling system is shown on Figure 4-4. Primary-chemical sludge was automatically blown down at various time frequencies from the solids-contact unit to a surge tank at a rate of about 200 gpm for about 8 seconds duration. A liquid level control operated Moyno Pump fed the gravity thickener at a rate of 3 to 5 gpm until the surge tank was emptied. In May 1973, a 3.5-ft diameter tank was placed inside the original 8-ft diameter thickener. Figure 4-5 is a picture of the surge tank-thickener arrangement.

Sample taps were located at 6-inch intervals for slurry pool level monitoring and sampling. The overflow from the thickener was passed through a coarse screen to a surge tank. From here the overflow was automatically pumped through a 1-inch totalizing water meter to the chemical treatment unit reaction zone liquid surface.

The thickener drive rotated the rakes and pickets at 16 rph. This resulted in a 6.7 and 2.9 fpm rake tip speed for the 8 and 3.5-ft diameter thickeners, respectively.

Originally, thickener blowdown was automated to coincide with blowdown from the chemical treatment unit. The duration of blowdown was adjusted by the operators to result in a desired volume of sludge. Because of the inability to maintain a desired near constant sludge pool depth in the 3.5-diameter thickener using the automated approach, manual blowdown was used.

Thickener underflow was routed, by gravity, to a 560-gal sludge storage and inventory tank. This vessel contained a radial flow turbine mixer to insure uniform suspension of the sludge prior to obtaining samples for laboratory analysis. The volume of sludge was determined by depth measurements. All chemical-primary sludge produced was collected, volume inventoried and samples obtained for solids analysis.

When vacuum filtration studies were programmed, the inventoried sludge provided the feed. When vacuum filtration was not programmed, the sludge was wasted to drain after volume inventorying and sampling. The vacuum filter used was the same unit used for the carbon regeneration system and will be described in a later section.

Chemical Storage and Handling

High purity alum and ferric chloride was purchased in dry form. Alum was obtained in 100-lb bags and the ferric chloride in 350-lb drums. These chemicals were intensely mixed

FIGURE 4-4:
SLUDGE TREATMENT SYSTEM

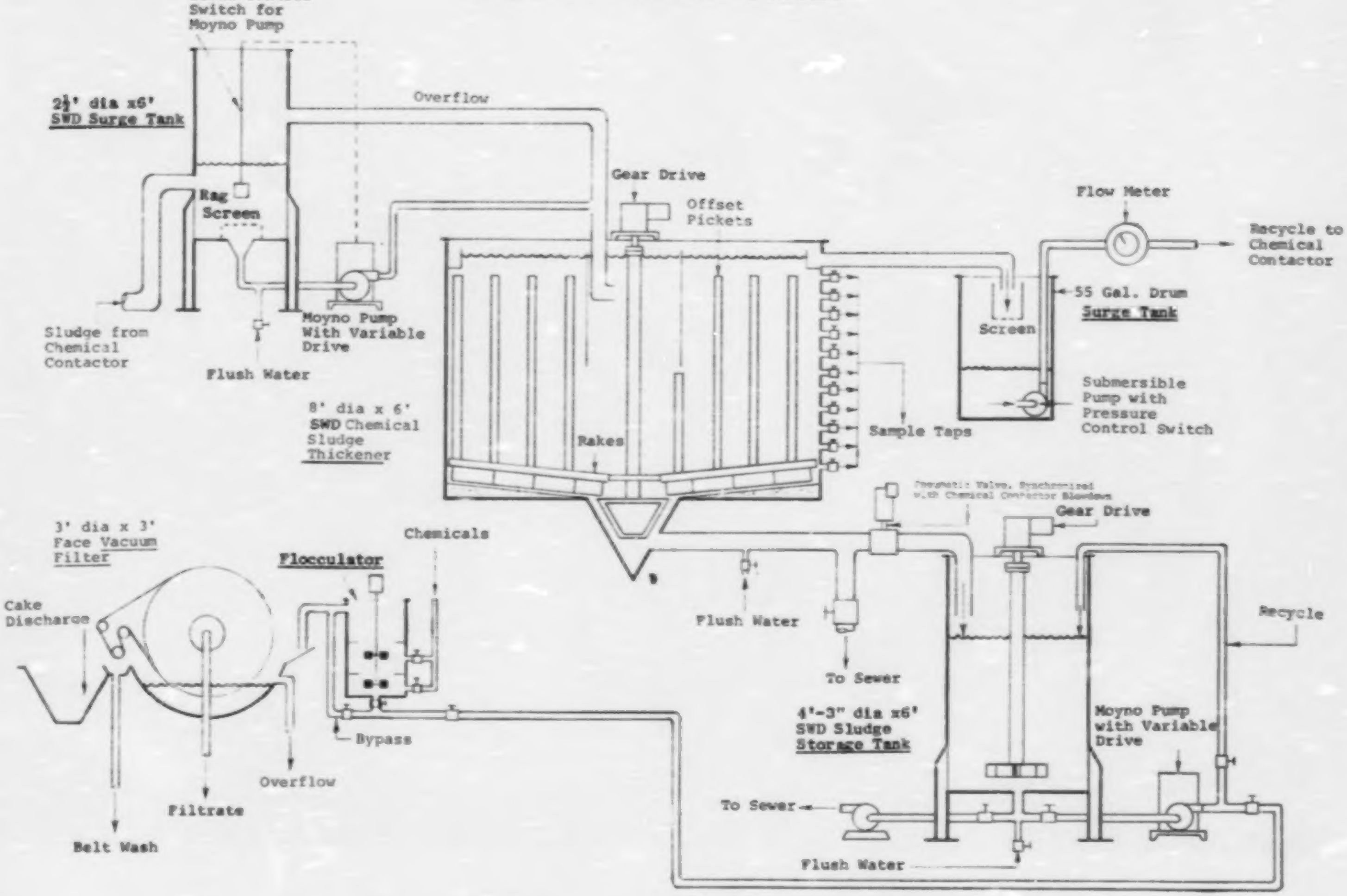
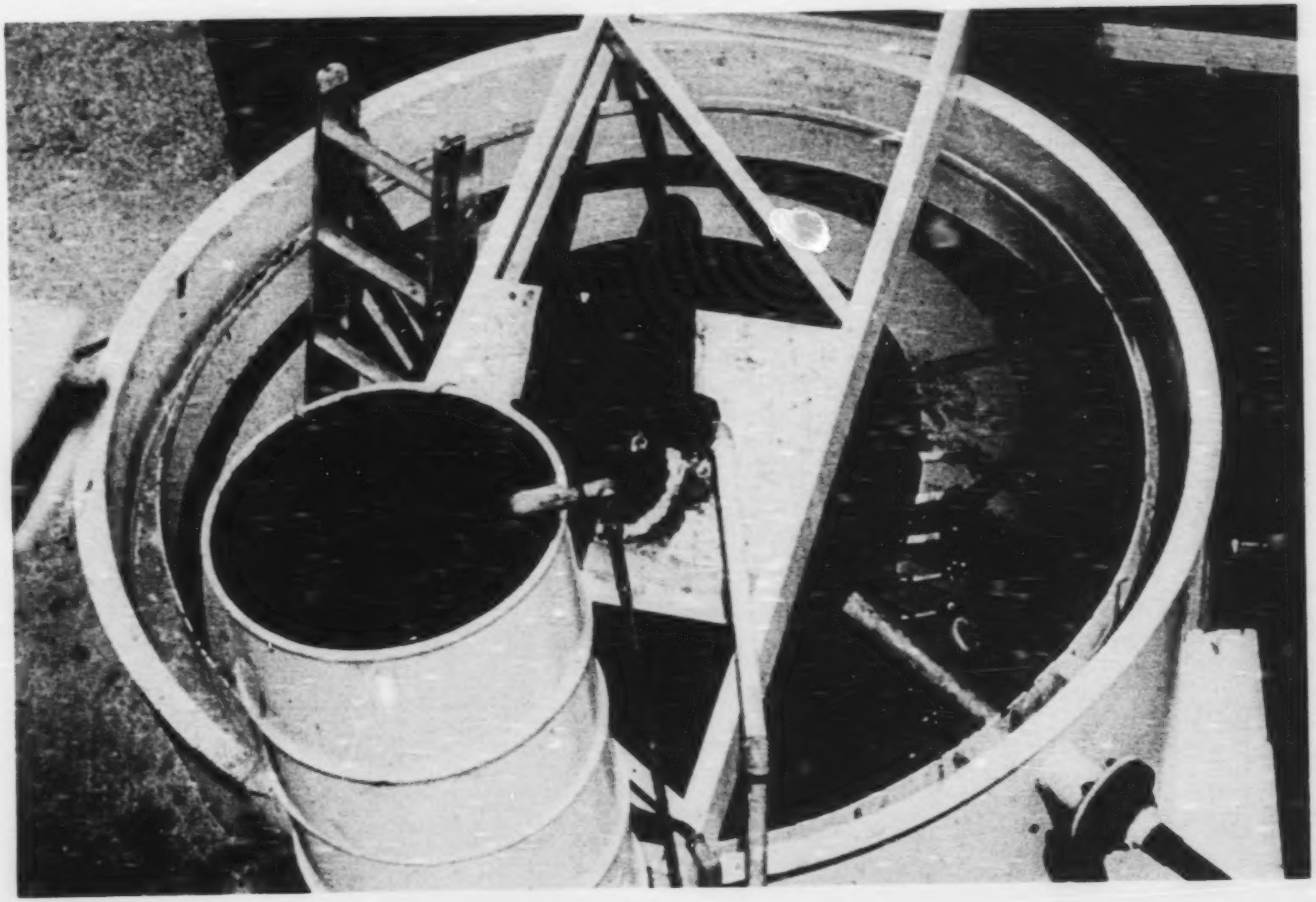

Level Control Switch for Moyno Pump
2½' dia x6' SWD Surge Tank
Overflow
Rag Screen
Sludge from Chemical Contactor
Flush Water
Moyno Pump With Variable Drive
8' dia x 6' SWD Chemical Sludge Thickener
Gear Drive
Offset Pickets
Rakes
Sample Taps
Flow Meter
Recycle to Chemical Contactor
55 Gal. Drum Surge Tank
Screen
Submersible Pump with Pressure Control Switch
Pneumatic Valve, Synchronized with Chemical Contactor Blowdown
Gear Drive
Flush Water
To Sewer
4'-3" dia x6' SWD Sludge Storage Tank
Recycle
Moyno Pump with Variable Drive
To Sewer
Flush Water
3' dia x 3' Face Vacuum Filter
Flocculator
Chemicals
Cake Discharge
Bypass
Overflow
Filtrate
Belt Wash

FIGURE 4-5: CHEMICAL SLUDGE SURGE TANK AND THICKENER



with tap water in a 55-gal drum to allow complete solution. The concentrated solution was then gravity fed to a 560-gal rubber lined tank where it was mixed with dilution water to achieve the desired feed concentration. The stock feed solution was fed to the flash mix tank by positive displacement metering pump through PVC piping.

Concentrated sulfuric acid was purchased in bulk quantities and stored in a 1000-gal covered mild steel tank. The acid was fed to the flash mix neutralization tank by a positive displacement metering pump through black iron piping. An automatic acid feed system consisted of a process pH meter, recorder and controller. The pH probe was located in a continuously flowing sample of neutralization tank overflow. This placement resulted in a control response lag time of less than 3 minutes. The controller had an adjustable low pH set point, below which the acid feed pump would turn off. As the pH increased above the set point, the acid feed pump would turn on. Manual trim of the acid pump rate allowed minimizing frequent OFF-ON pump action.

High purity hydrated lime ($Ca[OH]_2$) was obtained in 50-lb bags. Dry lime was continuously slurried with water and pumped through plastic tubing by a positive displacement pump to the chemical treatment unit draft tube. The lime slurry pump was operated at a constant rate sufficient to prevent deposition in the feed lines. Lime dosage was adjusted manually by a stroke adjustment on the dry lime feeder drive. During diurnal flow operation the dry lime feeder drive motor speed was automatically paced to plant flow.

A 560-gal (4.24-ft diameter) polymer make-up tank was provided for dissolving dry polymer into pilot plant effluent or tap water and storage of polymer solution during use. Polymers were dissolved by aspiration into the make-up water. High speed mixers (1750 rpm) were used to aid the solution of polymers and provide intermittent mixing of the tank contents. Polymer solution was fed by positive displacement metering pumps to the chemical treatment units.

All chemical feed pumps, carbon feed pumps and process stream samplers were driven by DC motors. The same voltage was applied to all motors and came from a single electronic controller. This controller received a signal via a pneumatic-electrical converter from the raw wastewater flow recorder-controller. Thus all reagent pump speeds were proportioned to raw wastewater flow.

CARBON TREATMENT SYSTEM

Carbon Contactors

Carbon treatment of chemically clarified raw wastewater was accomplished in one or two 10-ft diameter by 10.5-ft SWD solids-contact treatment units. These units were identical to the chemical treatment unit (Figure 4-3) except that they had no surface skimmers, flash mixers, or deflection plates, the turbines were not modified, and the effluent launders were of the submerged orifice type. The units were plumbed to be operated either in parallel, as single-stage carbon contactors, or in series, as a two-stage counter-current contacting system.

Carbon feed slurry (virgin or regenerated) was fed to the influent line of the carbon contactors by a positive displacement metering pump. When the two carbon contactors were operated in series, the virgin carbon slurry was fed only to the influent line of the second-stage unit. Underflow from the second-stage unit was intermittently pumped by a Moyno Pump to the influent line of the first-stage unit. Spent carbon was automatically blown down from the first-stage unit by gravity flow to the spent carbon slurry thickener (to be described later).

Carbon Make-Up and Feeding

Carbon was purchased in 35 or 50-lb bags. The type of virgin carbon used was Aqua-Nuchar.*

The virgin carbon make-up and feed tank is depicted on Figure 4-6. The 840-gal rubber lined and covered tank was outfitted with a two prop axial flow mixing impeller. During and just after addition of virgin carbon the air above the liquid was sucked through a water scrubber to remove carbon dust particles.

Routine carbon make-up procedure entailed first filling the tank about one-fourth full of tap water. The water door spray flow and vacuum pump were then started. Carbon was manually poured, very slowly, into the tank. After the desired weight of carbon had been added and wetted, tap water was added to a predetermined depth to provide a 5 percent by weight carbon feed slurry.

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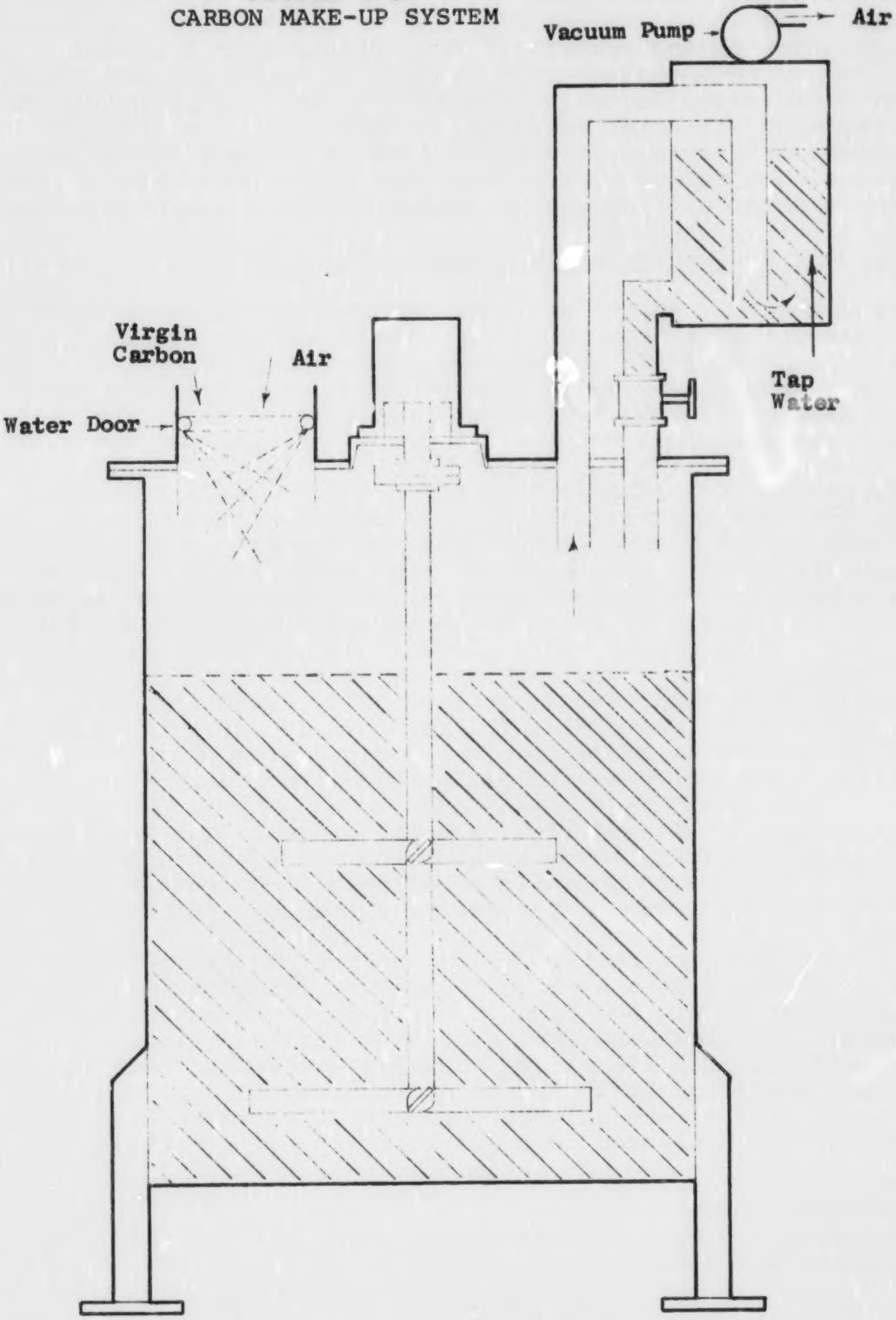
*A product of Westvaco, Covington, Virginia

FIGURE 4-6:
CARBON MAKE-UP SYSTEM


Vacuum Pump
Air
Virgin Carbon
Air
Water Door
Tap Water

Regenerated carbon slurry of about 15 to 20 percent total solids was stored in either a 560 or 840-gal tank containing axial flow mixing turbines. When required, a measured volume of regenerated carbon slurry of known concentration was transferred to the carbon make-up and feed tank and diluted with tap water to about 5 percent by weight slurry concentration. From the known total and fixed carbon solids values, the carbon slurry feed rate was established to provide a desired fixed carbon dosage to the carbon contactors.

Granular Media Filtration (GMF) Station

Effluent from one of the carbon contactors flowed by gravity to the granular media filter. The filter was 3.5-ft diameter by 6-ft high and of conventional downflow gravity design with a coal-sand filter bed. The specific details of the depth and size of media used will be presented in Section VI of this report.

Figure 4-7 is a schematic of the granular media filtration station. In the filtering mode, influent flows through a U-tube to the filter inlet. The dual-media filter bed rested directly on a false bottom of Pora-Septa* media. Flow was collected in the filter outlet plenum and routed to an effluent weir box located at an elevation which insured that liquid in the filter bed was never at less than atmospheric pressure. Filter flow rate was indicated by an orifice plate by-pass flow meter and an orifice plate and differential pressure cell arrangement for remote read-out. A sample of filter effluent was continuously pumped from the effluent weir box to a composite sampler and process turbidimeter.

Filtered water flowed from the effluent weir box to a 5000-gal (12-ft diameter) clearwell which provided treated water storage for plant use, filter backwash, and heat exchanger water for the carbon regeneration system.

A column of pressure taps spaced at 6-inch intervals was located on the filter housing. Pressure differentials between the outlet plenum and each pressure tap were automatically measured and recorded once each hour for a duration of 8 to 10 minutes. The filter was manually backwashed when the total filter headloss (including underdrain) reached a value of about 7 feet of water. The operational sequence during backwashing was presented in a previous report.[1] Backwash water

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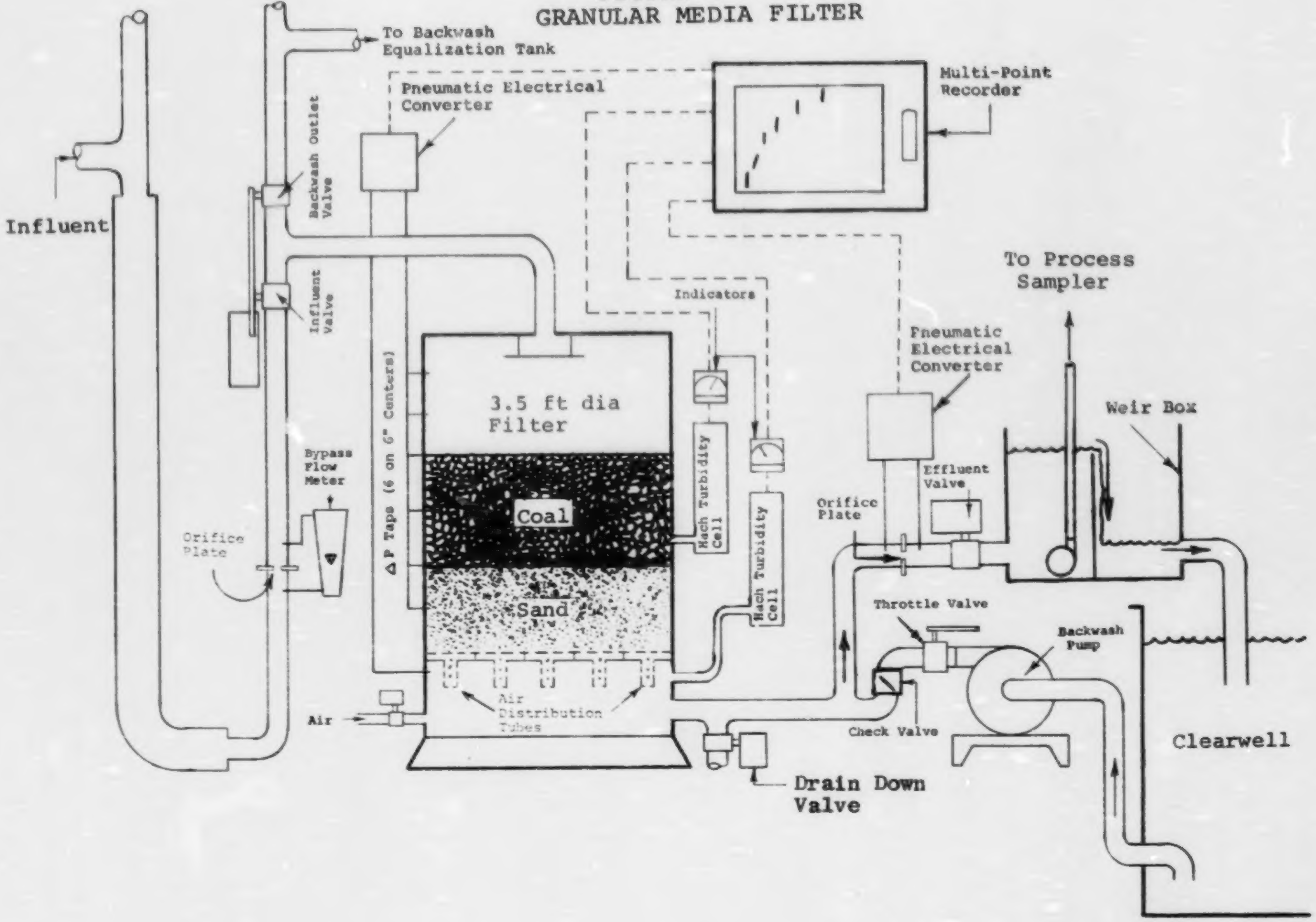
*Manufactured by Multi-Metal Wire Cloth Incorporated, Tappan, New York

FIGURE 4-7:
GRANULAR MEDIA FILTER


To Backwash
Equalization Tank
Pneumatic Electrical
Converter
Multi-Point
Recorder
Influent
Backwash Outlet
Valve
Influent
Valve
Indicators
To Process
Sampler
Pneumatic
Electrical
Converter
3.5 ft dia
Filter
Weir Box
Bypass
Flow
Meter
ΔP Taps (6 on 6" Centers)
Hach Turbidity
Cell
Hach Turbidity
Cell
Effluent
Valve
Orifice
Plate
Coal
Orifice
Plate
Sand
Throttle Valve
Backwash
Pump
Air
Air
Distribution
Tubes
Check Valve
Clearwell
Drain Down
Valve

was collected in a 3600-gal (10-ft diameter) equalization tank which was equipped with a low speed, radial flow, mixing turbine. Equalization tank contents were pumped to the reaction zone of the first-stage carbon-contactor at a rate of about 5 gpm.

Samples from the filter outlet plenum and just above the coal-sand interface were continuously passed through two Hach Low Range CR turbidimeters.* The electrical output signal from these turbidimeters, as well as flow orifice plate and pressure tap pneumatic-electrical converters were continuously fed to a multiple-point recorder for display and recording.

Carbon Regeneration System

The regeneration system consisted of four major unit operations: sludge concentration by gravity thickening, vacuum filter dewatering, regeneration in a fluidized bed furnace and off-gas carbon capture by wet venturi scrubbing. This system is shown on Figure 4-8.

The spent carbon gravity thickener was a 5-ft diameter by 5-ft SWD unit. Thickener underflow was transferred to a carbon inventory and sampling tank (about 500 gal). A small Moyno Pump was used for this transfer operation. The volume of thickened carbon slurry was measured. Then, after thorough mixing, a representative sample was obtained for laboratory tests. The inventoried, thickened carbon was then transferred to a 1700-gal conical storage tank. To prevent bridging in the conical storage tank, carbon was recycled externally at a rate of about 5 gpm from the bottom outlet to the top of the tank. The storage tank also contained four air nozzles to allow for air mixing.

Vacuum filter dewatering was accomplished with a 3-ft diameter by 3-ft face continuous belt filter.+ The vacuum filter station consisted of the filter, flocculation tank, vacuum pump and a single filtrate pump and receiver, all mounted on a platform. All components of the vacuum filter which contacted carbon slurry were made of corrosion resistant thermoplastic. A rubber lined chemical flocculation tank had two mixing compartments to achieve chemical dispersion and floc growth. A 55-gal drum, high speed mixer and variable speed positive displacement diaphragm chemical feed pump comprised the conditioning chemical make-up and feeding facility.

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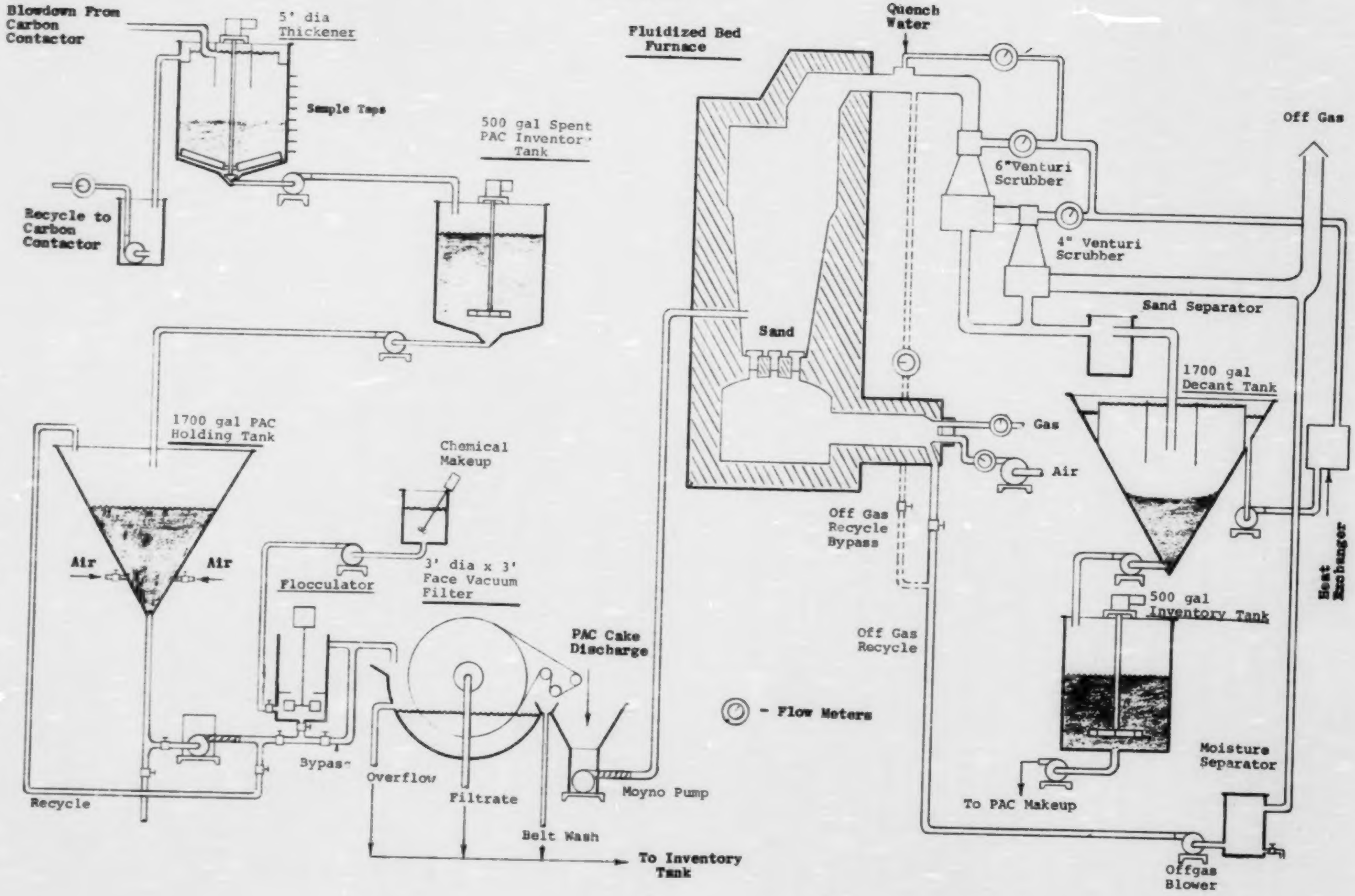
* Manufactured by Hack Chemical Company, Ames, Iowa

+ Manufactured by Eimco BSP Division of Envirotech Corporation, Salt Lake City, Utah

FIGURE 4-8:
POWDERED CARBON REGENERATION
SYSTEM


Blowdown From Carbon Contactor
5' dia Thickener
Sample Taps
Recycle to Carbon Contactor
500 gal Spent PAC Inventory Tank
1700 gal PAC Holding Tank
Air
Air
Chemical Makeup
Flocculator
3' dia x 3' Face Vacuum Filter
PAC Cake Discharge
Bypass
Overflow
Filtrate
Belt Wash
Recycle
Moyno Pump
To Inventory Tank
Fluidized Bed Furnace
Sand
Quench Water
6"Venturi Scrubber
4" Venturi Scrubber
Off Gas
Sand Separator
1700 gal Decant Tank
Gas
Air
Off Gas Recycle Bypass
Off Gas Recycle
- Flow Meters
500 gal Inventory Tank
Heat Exchanger
Moisture Separator
To PAC Makeup
Offgas Blower

The vacuum filter cake discharged into a hopper which had a capacity of about 300 lb of wet carbon filter cake. A ribbon screw advanced the cake along the bottom of the hopper to the throat of the furnace feed pump. The ribbon screw reconstituted the filter cake to a form of paste which the Moyno Pump readily transported to the fluidized bed furnace.

All carbon bearing streams (e.g., vacuum filter filtrate, belt wash, overflow, spillage and cleanup) were collected via a floor drain sump and pumped to a 3600-gal (10-ft diameter) inventory tank.

The outer dimensions of the carbon regeneration furnace were 5.25-ft diameter by 14-ft high.* The outer steel shell was lined with suitable fire and insulating brick. A major modification was made to furnace internal geometry mid-way in the contract period. An additional modification was made in the method used for controlling fire box temperature and fluidized bed O_2 levels. These changes will be discussed in Section VIII of this report. Thermocouples and pressure taps were used to monitor temperature and pressure profiles. Stainless steel tuyeres (nozzles) were provided to retain the sand and provide distribution of fire box gases into the sand bed.

The fuel used was natural gas. Combustion air was provided by a 200-SCFM capacity blower. Hot gas flow from the fire box at a superficial velocity of just over 1-ft per second was required to maintain the 15 by 30 U.S. Mesh sand bed in a fluidized state at operating temperatures.

The hot furnace off-gas, containing the regenerated carbon, was quenched to about 200°F by the addition of about 15 gpm of recycled water to the furnace outlet duct. The cooled gases were then passed through a 6 and then a 4-inch diameter wet venturi scrubber. These scrubbers, operated in series, captured essentially all of the regenerated carbon. Scrubber water flow was about 30 and 15 gpm to the 6 and 4-inch venturis, respectively. Scrubber water underflow was passed through a cyclone separator (a 55-gal drum) for sand and scale removal and then to a 1700-gal conical decant tank which provided some separation of carbon. Decant tank overflow was pumped through a heat exchanger prior to reuse as quenching and scrubber water. Pilot plant effluent was used for heat exchanger water. The fluidized bed furnace and venturi scrubbing system were highly instrumented for safety and ease of operation.

* Manufactured by BSP Division of Envirotech,
Brisbane, California

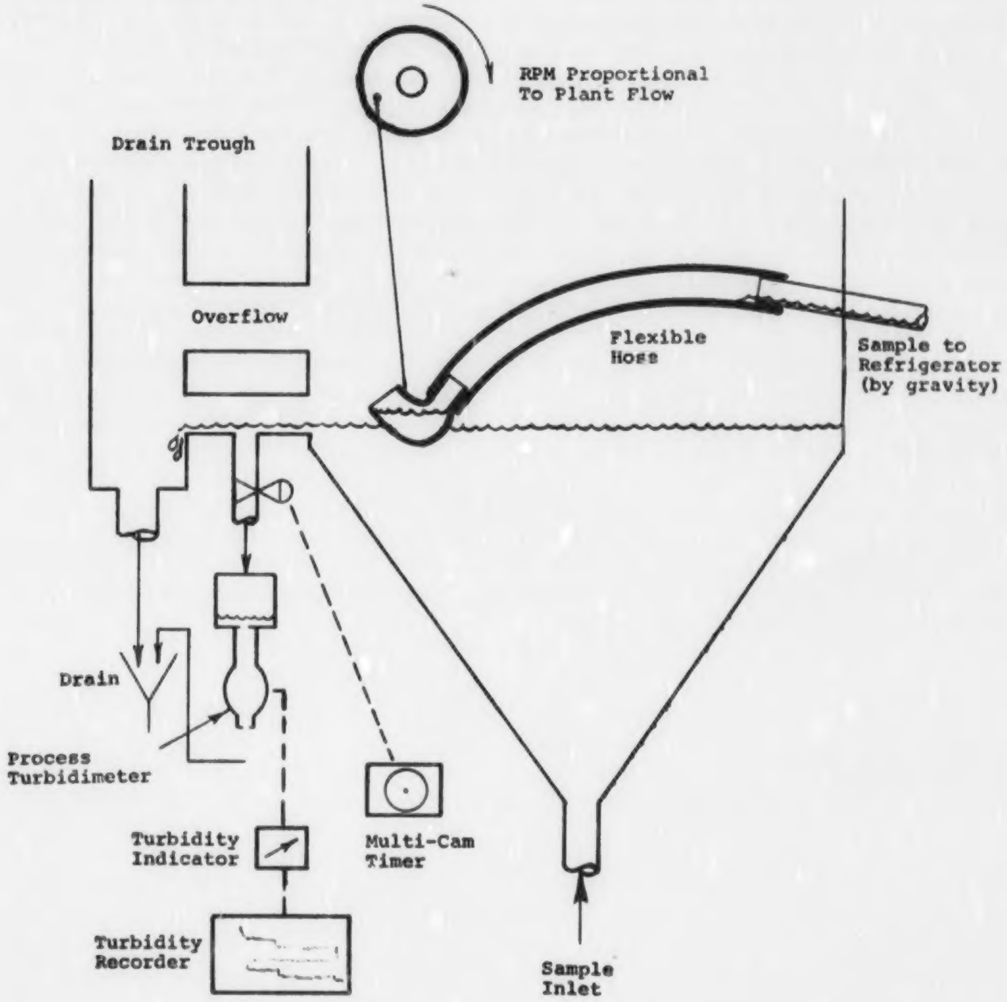
COMPOSITE SAMPLER AND PROCESS TURBIDIMETER

An automatic sample compositor, and a turbidity measuring and recording system was developed for this project. (See Figure 4-9.) All process stream samples (see Figure 4-2 for location) were piped to a central location. Each sample stream passed upward into a hopper shaped container, sized to maintain solids in suspension. Sample continuously overflowed the hopper via a 1-inch diameter horizontal pipe to a common waste drain. Normally closed solenoid valves were located in a vertical line just below the horizontal pipes. Outlets from all five solenoids emptied into a common trough which drained into a flow-through process turbidimeter. A multiple cam timer programmed each solenoid to be opened for 12 to 15 minute intervals once each hour. Twenty-four hour, flow proportioned composite samples were obtained from the hopper container mentioned previously. Approximately 50 mℓ of each sample was collected with a dip type sampler each 7 to 20 minutes. The actual frequence of sample pickup was directly proportional to raw wastewater flow. Samples were transferred to refrigerated storage vessels by gravity flow through plastic hoses. Composited samples were refrigerated at about 34°F until removed for analysis. The raw wastewater was sampled separately with a dip type sampler located at the raw wastewater sump.

LABORATORY FACILITIES

Analytical laboratory facilities were located at the pilot plant site. The laboratory was equipped to conduct all analysis reported except sulfides and dissolved metal analysis. These analyses were made at the Envirotech Research Center, Salt Lake City, Utah.

FIGURE 4-9:
COMPOSITE SAMPLING AND PROCESS
TURBIDITY MONITORING DEVICE


RPM Proportional
To Plant Flow
Drain Trough
Overflow
Flexible
Hose
Sample to
Refrigerator
(by gravity)
Drain
Process
Turbidimeter
Turbidity
Indicator
Multi-Cam
Timer
Turbidity
Recorder
Sample
Inlet

SECTION V

OPERATING PROCEDURES

The basic approach to pilot plant operation consisted of the following steps: establish desired operating conditions, operate the plant, collect operation data, collect samples, analyze samples, and reduce data. No step was initiated until the preceeding steps had been performed satisfactorily.

The desired operating conditions were established from the contract statement of work, feedback from reduced data, and/or operational planning. Some of these conditions were pretreatment chemical type and dosage, diurnal flow pattern, carbon dosage and number of carbon contacting stages, etc.

Plant operators would routinely monitor effluent clarity, chemical dosages, sludge profiles, underflow sludge blowdown rates, etc., to verify that specified operating conditions were being met. After these conditions were met, operating, inventory and mechanical data were observed and recorded. When these data substantiated that specific operating conditions were attained, sample collection was started.

The project chemist and plant operating engineer would determine daily if samples were representative, prior to laboratory analysis.

The operation data, inventory numbers, and laboratory results were then reduced. Actual reagent dosages, sludge production, pollutant removals, etc., were computed. Occasionally the need for additional data, special studies, and operational changes were identified by this routine data reduction.

CHEMICAL FEEDING

Reagent chemical stock flow to each unit was routinely measured (ℓ/min) and recorded about 6 times per day. In addition all reagent stock solutions or slurries were volume inventoried once each day.

Samples of alum and ferric chloride feed solutions were occasionally analyzed for aluminum or ferric iron to determine actual concentrations. These concentrations were compared with chemical makeup data to verify the reliability of the latter.

Available lime determinations were occasionally made when using lime. The total number of bags used and their contents weight was recorded when added to the dry lime feeder. Routine grab samples of the dry lime feed were weighed and the feed rate in g/min recorded.

The number of bags used for carbon makeup were recorded along with the manufacturer's listed bag weight (35 or 50 lb). On one occasion, a weight check indicated that the actual carbon weight was about 15 percent less than the manufacturer's listed weight for a given bag. Samples of the carbon feed slurry were occasionally grab sampled and suspended solids determined to verify the computed feed slurry concentration based on makeup data. When using regenerated carbon the feed slurry was made up from concentrated slurries of stored regenerated carbon by dilution to the desired feed strength.

SOLIDS-CONTACT UNITS

Instantaneous flow to the chemical treatment unit was automatically measured and displayed on a recorder. Flow to subsequent units was determined by subtracting routinely measured process sample flows from the displayed plant influent flow.

The approach to operation of the solids-contact treatment units was presented in a previous report.[1] A brief reiteration follows. It is important to keep in mind that the type of unit used could be operated either as a solids-contact or a solids-contact and sludge blanket clarifier. Sludge blanket clarification is defined as maintenance of at least two or more feet of slurry in the clarification zone. With reference to Figure 4-3, this means a slurry height at least two feet above the lower extremity of the reaction zone skirt. During this current study no sludge-blanket operation was specified, only solids-contact treatment - with gravity sedimentation.

Operation of the solids-contact units basically involved routine monitoring of sludge concentration profiles within the three units. Two operational specifications were given to the operators. The first was to maintain an approximate specified solids concentration within the reaction zone.

The second was to keep the top of the sludge slurry at or slightly below the lower extremity of the reaction zone skirt. An indirect method for determining sludge concentration was used. This method consisted of collecting two liters of sludge in a 2-liter graduated cylinder. The sludge was allowed to subside quiescently for either 5 or 10 minutes. The time was chosen such that the slurry interface was subsiding at a rate of less than about 1-inch per minute. For the carbon and lime sludges, 5 minutes were adequate. For the alum and ferric sludges, 10 minutes were required. At the desired time interval, the volume of settled sludge was observed and the percentage of the original sample volume recorded. This measurement was called the 5 or 10 minute settled sludge volume.

Approximately six times per day (twice per shift), the operators would determine settled sludge volume for samples from all three solids-contact units. These samples were taken at tap locations G and T on Figure 4-3. The values obtained were compared with the previously mentioned two operational specifications. If the slurry was at an undesirable high level, the rate of sludge withdrawal (blowdown) was increased. If the slurry level was too low, the rate of sludge withdrawal was reduced or temporarily shut off.

Infrequently, the pumping turbine and sludge rake rotational speeds were adjusted by the operators, if too high or too low a sludge concentration was observed for the reaction zone samples. Adjustments in turbine and rake speeds were very infrequent, sometimes not being required for 30 to 60 days. Sludge blowdown rate adjustments were more frequently required, sometimes being daily and other times being only weekly or monthly.

GRANULAR MEDIA FILTER

As indicated in Section IV, the granular media filter (GMF) was highly automated for backwashing and monitoring of performance (e.g., flow rate, headloss and turbidity). The approach to operation of the filter station was mainly one of visual observations for operational problems, calibration of sensors and determining the need for backwashing.

About six times per day operators observed filter flow rate and headloss and analyzed effluent grab samples for turbidity and light transmittance. The operator determined the need for backwashing by observation of either a total bed headloss greater than about 85 inches of water or break-through of carbon fines. Penetration of carbon into the sand layer was visually observed through a plexiglass window. Carbon fine penetration into the sand coupled with an effluent quality deterioration trend, observed by continuous turbidity monitoring, indicated break-through. After manual initiation of the automatic backwash controller, the operator usually observed the filter backwash sequence. Rates and durations of initial water level drain down, air and water flows were frequently noted. Filter bed expansion during water backwash was routinely noted. Abnormalities in the "quality of media fluidization" or the presence of any "mud balls" were also noted.

Just prior to and shortly after backwashing the incremental headloss profile through the filter bed was frequently observed by the operator. A water manometer was used. Headloss profile at the end of a filter cycle indicated the solids removal patterns. Headloss profile at the start of the filter run was used to verify that effective backwashing was routinely being obtained.

GRAVITY SLUDGE THICKENERS

Operation of the chemical-primary and spent carbon sludge thickeners consisted of routine monitoring of sludge pool depths. Depths were denoted as being the highest sample tap at which copious amounts of sludge was observed. The operators were instructed to maintain the sludge level between two specified sample taps. This was accomplished by manual operation of thickener blowdown.

At about 4-hr intervals, 200-mℓ grab samples of clarifier underflow sludge and thickener overflow were obtained and composited over a 24-hr period. Suspended solids analysis of these samples indicated thickener solids captures.

When specified, 5 gal of 4-hr grab composite samples of clarifier underflow sludge were obtained by the operators. These samples were used to perform laboratory thickening tests. Usually, on the same day, a grab sample from each pilot plant thickener sample tap was obtained and total suspended solids determinations conducted. This data constituted the thickener solids profile (concentration versus depth).

VACUUM FILTER

Prior to the start of pilot plant vacuum filter operation for each type of sludge studied, filter leaf tests were run on composite sludge samples of gravity thickener underflow. A 0.1 sq ft filter leaf was used to determine the necessary type and dosage of chemical conditioners and an appropriate filter media type.

Just before the start of a vacuum filter run, suspended solids of stored and mixed gravity thickener underflow was determined. Chemicals were made up, and dosage and flow requirements determined. The sludge was then pumped approximately 40 ft to the flocculator where chemicals were added to and mixed into the sludge.

In order to maintain a constant drum submergence, and thus constant form and dry times, sludge was added to the vacuum filter faster than it could be dewatered. The excess sludge overflowed to recycle or waste. Drum rotation speed was started at the slowest possible level to achieve the thickest possible cake. The speed was then increased incrementally, taking data at each speed, until the cake was to thin to discharge. At this point, the filter run was concluded. This approach resulted in identification of the maximum possible yield for a dischargeable filter cake.

At each vacuum filter drum speed, the following samples were collected: composited samples of sludge feed before and after chemical addition, cake samples using a "cookie cutter" having an area of 0.085 sq ft, and a composited sample of filtrate. The difference between the suspended solids in the feed before and after chemical conditioning indicated solids increase due to the addition of chemicals. The difference between the suspended solids in the feed sludge before chemical conditioning and the filtrate indicated vacuum filter solids capture. Solids analysis of the filter cake samples, taken by the "cookie cutter", were used to compute vacuum filter yield and cake moisture content.

FLUIDIZED BED FURNACE

The operation of the carbon regeneration system was on a batch basis. Carbon losses were determined by conducting a solids balance across the system. The quantity and characteristics of spent carbon were determined prior to regeneration run. All carbon not reaching the furnace was collected and quantified. The difference was the carbon fed to the furnace. Recovered carbon was collected, and the quantity and characteristics determined. Carbon losses were based on fixed carbon analysis. The characteristics of regenerated carbon were measured by iodine number, molasses number, and equilibrium adsorption isotherm tests and by reuse in the pilot plant.

Safety procedures dictated the following start-up sequence be followed for the fluidized bed furnace: scrubber water on, burner air flow on, pilot gas flow on, pilot ignition established, and finally burner gas flow on. The furnace was then brought up to operating temperature very slowly. This normally took 3 to 4 days when starting at ambient temperature.

To avoid carbon combustion in a fluidized bed furnace, there should be less than about 1 percent oxygen by volume in the fluidizing gas in the bed. This can be accomplished by burning air and gas in the firebox stoichiometrically. However, construction materials are not available to handle the resulting 3200°F temperature. Temperature in the firebox can be controlled by using excess combustion air. Excess air for cooling results in excess oxygen in the fluidized sand bed where spent carbon is added. The original designers of the fluidized bed unit felt that gas could be injected into the bottom of the sand bed to scavenge the excess oxygen so that carbon would not be burned. This principal of operation will be referred to as bed injection gas (BIG). Temperature control can also be obtained by burning gas and air stoichiometrically and using an additional gas which contains little or no oxygen for cooling. These were the two methods of firebox temperature control used in this study.

When bed injection gas was used the following procedure was followed to ready the furnace for operation. The air and gas flow rates were set to provide the desired fluidization velocity. Proportions of these flows were determined from previous experience. Burner gas was then adjusted to provide a hot gas flow to the bed with an oxygen content of less than 1 percent by volume. The off-gas recycle flow was trimmed to maintain a firebox temperature under 1090°C, yet maintain bed fluidization.

Water was injected into the bed prior to feeding carbon in order to hold the bed temperature to within about 100°C of the desired operating temperature. This water injection was used during final trimming of the oxygen which was measured in the stack gas.

The vacuum filter capacity was about four times greater than the furnace capacity and, therefore, was run first to fill the feed hopper and then as needed. When the oxygen had been trimmed and the feed hopper was filling, the feed to the furnace was started.

The feed pump to the furnace was automatically controlled to maintain a set bed temperature. Minimal operator attention was needed during the run because of the automatic features of the furnace control. These features would shut the furnace down and/or sound an alarm to alert the operator if any abnormal condition occurred. Some of these conditions were: low scrubber water flow, low burner air flow, flame out or high bed temperature.

During the course of a regeneration run, pressures and temperatures at various points through the furnace were recorded (i.e., firebox, bottom of bed, upper bed, free-board). The stack gas was routinely analyzed for oxygen content. The flow through the heat exchanger was regulated to maintain a scrubber water temperature of 40 to 60°C.

At the conclusion of a regeneration run, the carbon cake feed was allowed to run out and the feed hopper, pump and piping was flushed with water. The furnace was shut down according to the manufacturer's instructions. About 3 to 4 days were required to cool the furnace and sand bed to a temperature approaching ambient.

Collection of regenerated carbon was accomplished by the following technique. Carbon in the scrubber water decant tank was allowed to settle and then clean supernatant decanted. Water was run through the scrubber system flushing out all deposited carbon to the decant tank. All deposition points (i.e., sand separator, launders, etc.) were flushed with water and all carbon returned to the decant tank. The carbon in the decant tank was again allowed to settle and clear supernatant again decanted. The remaining carbon was transferred to an inventory tank where it was mixed, volume inventoried and sampled. The regenerated carbon was then pumped to the carbon makeup and feeding tank as needed.

LABORATORY OPERATIONS

The spectrum of analytical tests routinely run on automatically composited process stream samples is shown in Table 5-1. Sampling and laboratory test procedures are presented in Appendix A. These routine analyses were run 4 days per week. In addition to process stream samples, daily grab composites of all solids-contact reaction zone and blowdown sludge samples, and thickener overflows required suspended solids analysis. All chemical stock feeds, and chemical-primary and spent carbon thickener blowdowns were inventoried, sampled and appropriate mineral or suspended solids determinations made when scheduled.

Operation of the vacuum filter and the carbon regeneration system were on an intermittent basis. Each of these operations generated about 20 to 40 samples for suspended solids and, for a few samples, mineral analysis. Spent and regenerated carbon samples were analyzed for volatiles, ash and fixed carbon solids and other adsorptive characteristics.

TABLE 5-1: SCHEDULED ANALYTICAL TESTS OF PROCESS STREAM COMPOSITE SAMPLES

Water Quality Parameter*	Raw Wastewater	Chemical Treatment	Carbon Contactor First Stage	Carbon Contactor Second Stage	GM Filter
Turbidity	a	a	a	a	a
Suspended Solids	a	a	a	a	a
Light Transmittance	-	a	a	a	a
Total & Soluble Phosphorus	a	a	-	-	a
Total & Soluble Ortho-Phosphate	a	a	-	-	a
Chemical Oxygen Demand	a	a	-	-	a
Soluble Chemical Oxygen Demand	a	a	a	a	a
Biochemical Oxygen Demand	b	b	-	-	b
Soluble Biochemical Oxygen Demand	-	-	b	b	-
pH	a	a	a	a	a
Alkalinity	d	d	-	-	d
Total Hardness	d	d	-	-	d
Calcium Hardness	d	d	-	-	d
Sulfides	b	b	b	b	b
Total & Soluble Iron or Aluminum	c	c	c	c	c

a - Routine analysis on 24 or 48 hour composites
b - No more frequently than once per week
c - Intermittant during $FeCl_3$ or alum campaign
d - Twice per week during lime campaign
* - Soluble pollutants were defined as those passing a 0.45 micron membrane filter

The laboratory personnel had to handle these and other special laboratory tests as time was available.

Laboratory gravity thickening and vacuum filtration leaf tests were conducted by plant operators and/or project engineers.

SECTION VI

PILOT PLANT PERFORMANCE

INTRODUCTION

The purpose of this section is to present and discuss the liquid treatment performance of the PAC-PCT pilot plant. The liquid phase of the plant includes the unit operations of chemical treatment, carbon contacting and granular media filtration (see Figure 4-2). Pilot plant effluent quality and consistency will be emphasized. Detailed discussions of sludge treatment, carbon regeneration and carbon system response are presented in subsequent Sections.

Pilot plant performance for various operating conditions will be presented and discussed. The data presented represents all available performance data except for some data collected in November and December 1972. During November and December 1972, a special study to identify the effect of anaerobic biological activity on soluble organic removal in the carbon contactors was conducted. The two carbon contactors were run in parallel, one with and the other without carbon feed. Results are presented in Section IX.

The effect of diurnal flow variations and operation on the performance of the solids-contact units will be discussed relative to effluent quality, sludge depth and solids concentration profiles, and down time due to treatment upsets.

The performance of granular media filtration will be discussed relative to the effects of bed design and operating conditions on suspended solids removal efficiency and patterns.

It should be recalled that an objective of this study was to evaluate the effectiveness of various filter bed designs. During the first half of the study, a very coarse coal media was used. This filter bed produced less than satisfactory suspended solids removal which affected pilot plant effluent quality.

PILOT PLANT PERFORMANCE

Figure 6-1 is a chronology of plant operation indicating major treatment variables and unit operations. Six operating periods were identified on the basis of uniform chemical treatment, flow variation (i.e., diurnal pattern) and/or carbon contactor operation (i.e., number of stages and carbon dosage). The operating periods will be presented chronologically. Average operating conditions, and influent and effluent quality, were determined for each period. Pilot plant effluent history curves were plotted to show the consistency of treatment achieved.

Operating Period I

Operating Period I was from July 13 to September 30, 1972, during Phase I of the contract work. The plant was operated at constant flow with sufficient hydrated lime added to achieve an operating treatment pH of 10.8 to 11.2. Constant flow was used because diurnal flow equipment was not yet installed. High carbon dosages and parallel single-stage carbon contacting were used in order to produce the maximum possible amount of spent carbon for regeneration studies. Development of the carbon regeneration system was a major objective of Phase I studies. The effluent from only one of the carbon contactors was filtered through a "course bed" granular media filter.

Table 6-1 presents average pertinent plant operating conditions and raw wastewater, chemical contactor, and plant effluent water quality. Figure 6-2 is a plant effluent history curve showing suspended solids, total and soluble COD and total phosphorus.

The granular media filter was operated intermittently from August 8 to September 17 because of new instrumentation installation, and electrical and mechanical problems. Because of these problems, filtered plant effluent data was not available for the entire operating period. Hence, plant effluent total COD is not plotted for the entire period because total COD analysis was not performed on carbon contactor effluents due to the presence of carbon suspended solids which contribute to COD.

Upon start-up of a chemical treatment solids-contactor, it is sometimes difficult to keep the dilute suspensions of fresh floc out of the clarification zone. After experiencing this problem, the flow was reduced by half to allow the chemical treatment unit to build a sludge

FIGURE 6-1:
CHRONOLOGY OF PLANT OPERATION

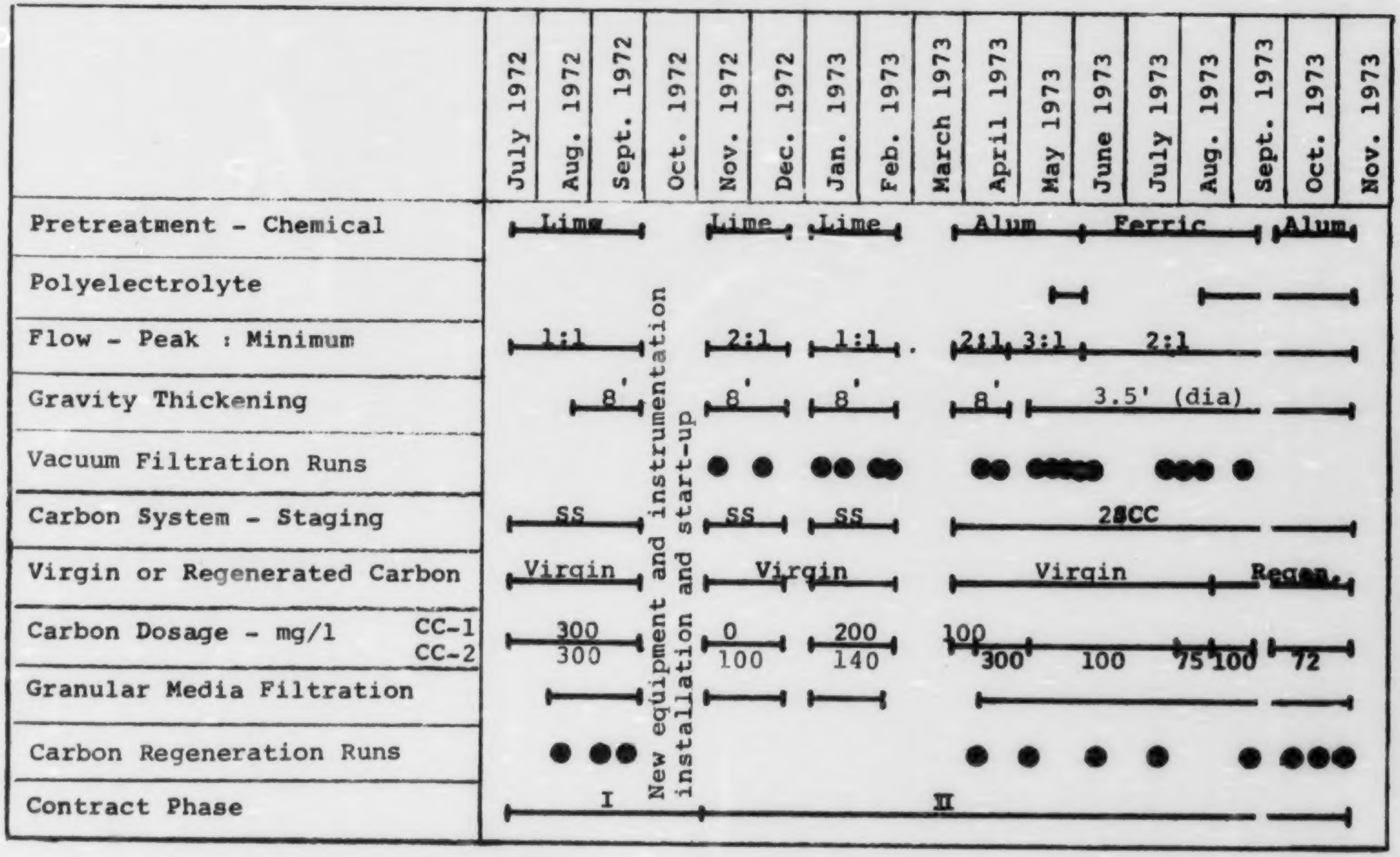


TABLE 6-1: OPERATING PERIOD I: AVERAGE OPERATING CONDITIONS AND TREATMENT RESULTS JULY 13 TO SEPTEMBER 30, 1972

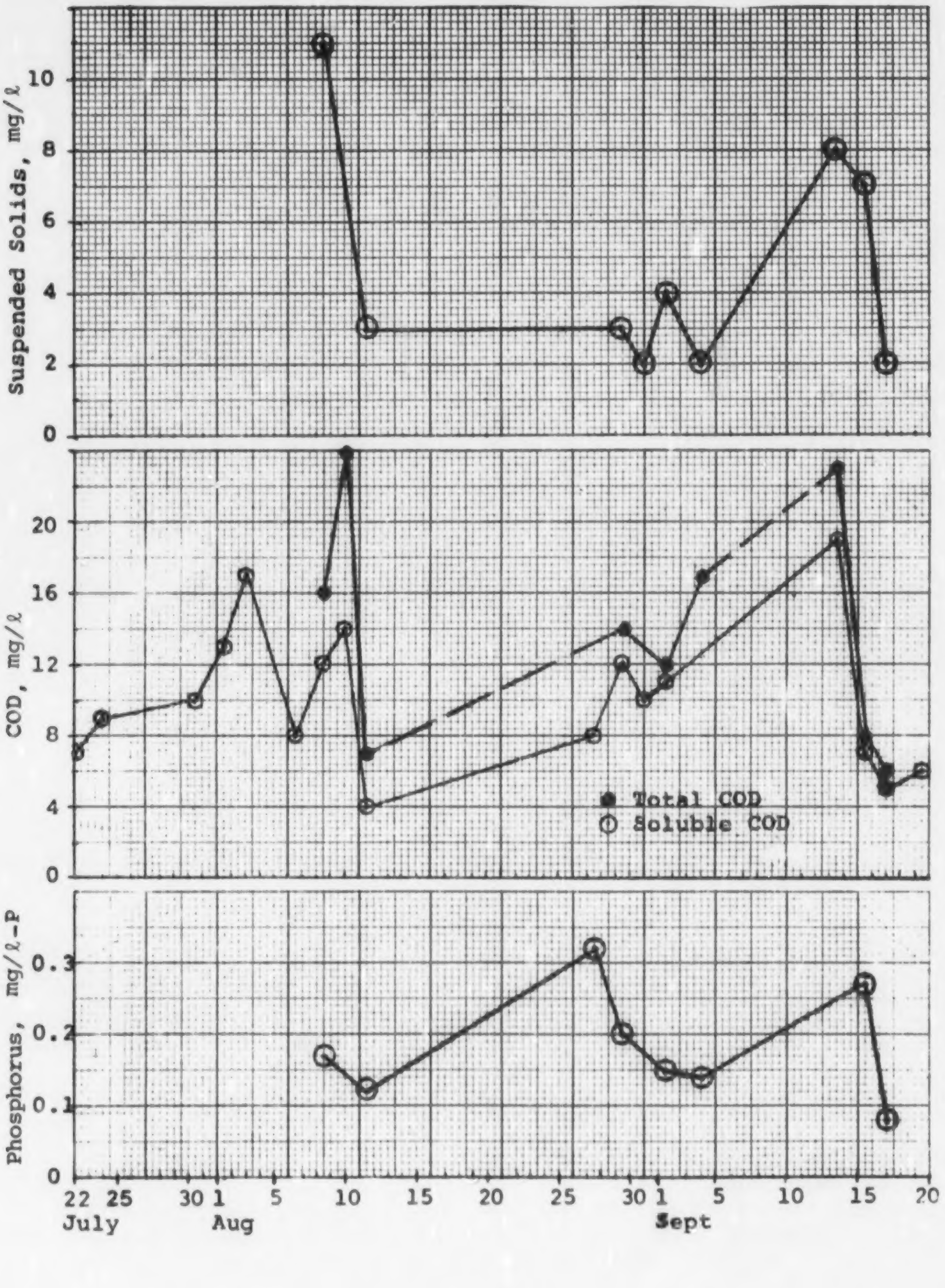
	Operating Conditions	
Chemicals: Hydrated Lime:	@ 460 mg/ℓ (10.8 to 11.2 treatment pH)	
H_2SO_4:	@ 2030 lb/MG = 244 mg/ℓ	
Average Hydraulic Loading:	Chemical Contactor	0.94 (0.49[a]) gpm/sq ft
	Carbon Contactor	0.66 (0.32[a]) gpm/sq ft
	GM Filter	2.1 gpm/sq ft
Carbon Contacting Mode:	Single stage	
Carbon Dosage:	First Contactor	280 mg/ℓ
	Second Contactor	300 mg/ℓ

Average Process Treatment Results	Raw Wastewater	Chemical Effluent	Neutralized Chemical Effluent	Plant Effluent
Turbidity, JTU	33	4	4	3
Suspended Solids, mg/ℓ	87	14	10	5
% Transmittance	-	-	-	97
Total P, mg/ℓ P	4.50	0.29	-	0.20
Soluble Total P, mg/ℓ P	3.16	0.14	0.15	0.11
Total σPO4, mg/ℓ P	2.82	0.10	0.06	0.11
Soluble, σPO4, mg/ℓ P	2.25	0.04	0.05	0.08
COD, mg/ℓ	136	-	55	14
SCOD, mg/ℓ	30	-	35	11
pH, units	7.5	10.2	7.1	7.2
Methyl Orange Alkalinity[b]	293	116	65	58
Phenolphtalein Alkalinity[b]	-	66	-	-
Total Hardness[b]	417	239	317	328
Calcium Hardness[b]	255	197	284	250

a - 7/29 to 8/9 only

b - mg/ℓ as $CaCO_3$

FIGURE 6-2: EFFLUENT HISTORY CURVE FOR OPERATING PERIOD I
(JULY 13 - SEPTEMBER 30, 1972)



inventory. This took several days and is indicated as the lower hydraulic loadings presented in Table 6-1.

The neutralized chemical treatment contactor effluent was not analyzed for total phosphorus because it was equal to that of the chemical treatment contactor effluent. The pH values shown in Table 6-1 are for stored composite samples and not instantaneous measurements. The 10.2 pH value reported for the chemical treatment contactor effluent is 0.6 to 1.0 unit below the actual operating pH.

A slight solubilization of COD was experienced across the chemical treatment step (i.e., increase in SCOD). The carbon contactors removed about 72 percent of the SCOD fed to them. Carbon contactor effluent suspended solids averaged 32 to 36 mg/ℓ, most of which was carbon.

During this operating period, the pilot plant removed 94 percent suspended solids, 96 percent total phosphorus, 90 percent total COD, and 21 percent hardness. Water stability analysis were made in an attempt to quantify the very stable character of several grab samples of chemical contactor effluent. The "Marble Test" was used. This test determines the difference, in pH units, between the observed pH and the pH of stability (pHs). If the difference is positive there is a depositing tendency and if the difference is negative (observed pH less than pHs) there is a tendency of dissolving calcium carbonate depositions.

Analysis of five grab samples collected on three different days indicated a range of Marble Test stability of from minus 0.08 to plus 0.02 pH units with an overall average of 0.00 pH units.

After 3 months of continuous operation using lime, no significant mineral deposition had occurred on the chemical contactor tank walls or launders. Some deposition of calcium carbonate was found on the inside of the reaction zone skirt near the issue of the pumping turbine. This deposition caused no operational or mechanical problems.

Generally, the average effluent qualities were considered good. However, the poor granular media filter performance caused considerable variability in the effluent quality.

Operating Period II

The second pilot plant operating period was from January 6 to February 20, 1973. The pilot plant was operated at constant flow due to instrumentation failure. New equipment was installed and debugged in October, 1972. As mentioned before, special studies were conducted in November and December, 1972. This accounts for the time lapse between operating Period I and II.

During Period II, sufficient hydrated lime was added to produce an operating treatment pH of 10.8 to 11.2. Parallel single-stage carbon contacting was used. Effluent from one of the carbon contactors was filtered through a "coarse bed" granular media filter.

Table 6-2 presents pertinent operating conditions and a tabulation of average plant influent and effluent and chemical contactor effluent water quality. As indicated in Table 6-2, different carbon dosages were used in the two carbon contactors. Figure 6-3 is the plant effluent history curve of this operating period. Due to operational problems and filter failure, the filter was only operated about 1/3 of this operating period. However, even when the filter was on stream it did a poor job of filtration. This can be seen in Figure 6-3 by considering the high filter effluent suspended solids and the large difference between total and soluble COD. Soluble COD removal was generally very good and consistent, whereas particulate removal was poor and variable.

During this period neither solubilization nor net removal of SCOD was experienced across the chemical treatment step. The carbon contactors removed about 72 percent SCOD and overflowed 20 to 30 mg/ℓ suspended solids.

The granular media filter removed only 33 percent of the suspended solids fed to it. Because of this poor suspended solids removal, the total COD of the plant effluent averaged 41 mg/ℓ. From Table 6-2 it is seen that 31 mg/ℓ of the total COD in the plant effluent was caused by suspended solids passing through the filter.

Because of poor granular media filter performance, the plant was only able to remove 81 percent suspended solids and 68 percent total COD. Had 95 percent suspended solids removal been achieved, 88 percent removal of COD would have been experienced. However, 96 percent removal of total phosphorus was achieved even with poor filter performance.

As in the previous operating period the wastewater was again slightly softened by solids-contact lime treatment as indicated by a 26 percent reduction in total hardness. Marble stability test results again showed that a very stable effluent was produced by the solids-contact chemical treatment unit. The average difference between the observed pH and

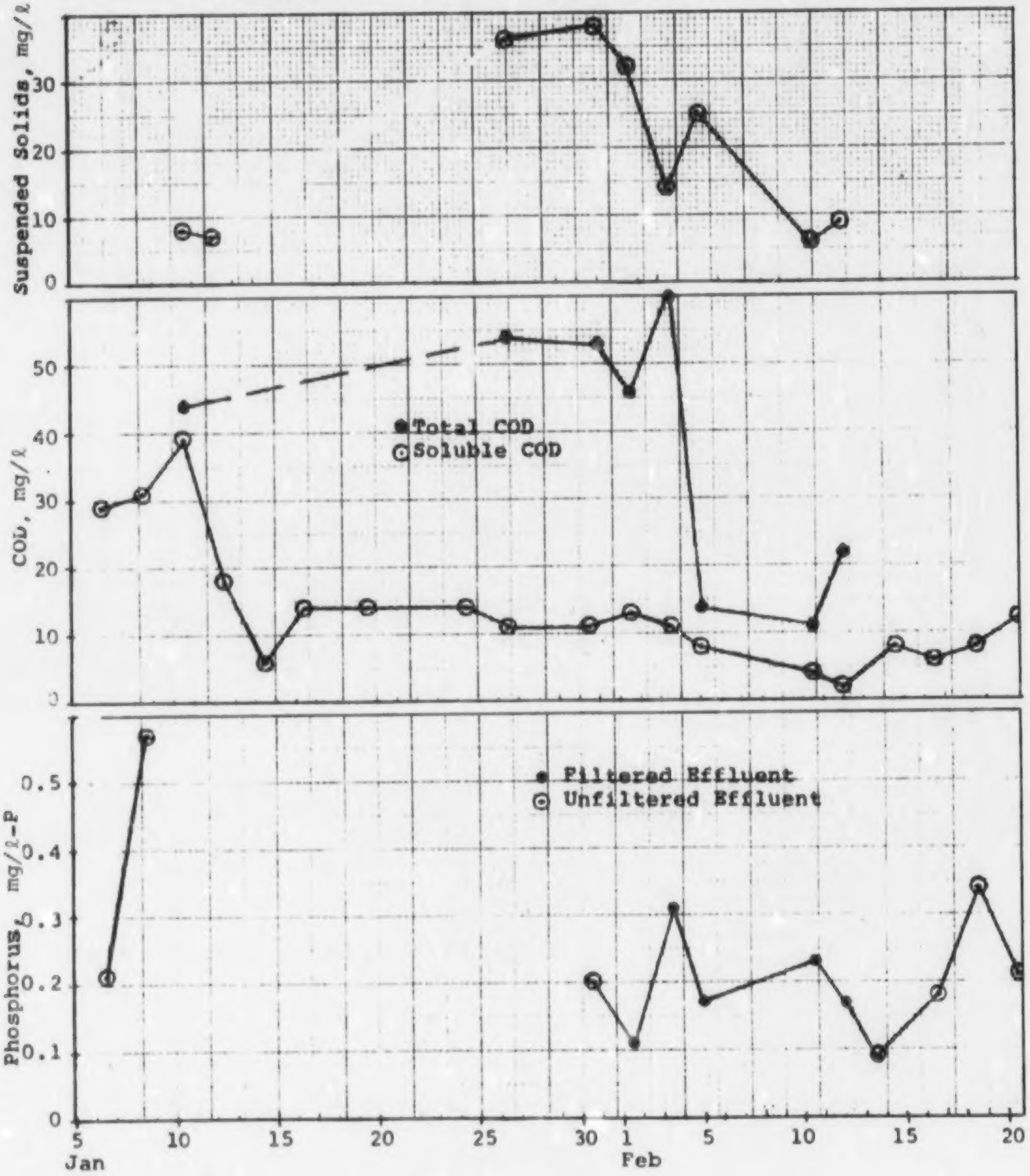
TABLE 6-2: OPERATING PERIOD II: AVERAGE OPERATING CONDITIONS AND TREATMENT RESULTS JANUARY 6 TO FEBRUARY 20, 1973

	Operating Conditions	
Chemicals:	Hydrated Lime	@ 490 mg/ℓ (10.8 to 11.2 Operating pH)
	H_2SO_4	@ 2080 lb/MG = 250 mg/ℓ
Average Hydraulic Loading: (constant flow)	Chemical Contactor	0.84 gpm/sq ft
	Carbon Contactor	0.56 gpm/sq ft
	GM Filter	3.0 gpm/sq ft
Carbon Contacting Mode:	Single Stage	
Carbon Dosage:	First Contactor	183 mg/ℓ
	Second Contactor	134 mg/ℓ

Average Process Treatment Results	Raw Wastewater	Chemical Effluent	Neutralized Chemical Effluent	Plant Effluent
Turbidity, JTU	47	7	3	4
Suspended Solids, mg/ℓ	99	27	9	19
% Transmittance	--	--	-	90
Total P, mg/ℓ P	4.88	0.26	-	0.21
Soluble Total P, mg/ℓ P	3.62	0.13	0.21	0.13
Total σ PO_4, mg/ℓ P	2.54	0.09	0.13	0.09
Soluble σ PO_4, mg/ℓ P	2.07	0.05	0.08	0.06
COD, mg/ℓ	127	--	46	41
SCOD, mg/ℓ	41	--	41	10
BOD, mg/ℓ	84	--	25	5
pH, units	7.4	10.3	8.0	8.0
Methyl Orange Alkalinity,[a]	342	142	64	49
Phenolphthalein Alkalinity,[a]	--	86	--	--
Total Hardness,[a]	484	251	387	360
Calcium Hardness,[a]	272	211	299	298

[a] mg/ℓ as $CaCO_3$

FIGURE 6-3: EFFLUENT HISTORY CURVE FOR OPERATING PERIOD II
(JANUARY 6 - FEBRUARY 20, 1973)



pH at stability was 0.01 pH units, with a range of only 0.01 pH units. These results are for eight grab samples collected of six different days.

Except for the presence of excessive carbon suspended solids, the quality of the plant effluent was considered very good. As seen in Figure 6-3, about 1 week after plant start, a very consistent plant effluent SCOD was produced.

Operating Period III

Operating Period III was from March 28 to June 7, 1973. The plant was operated with alum pretreatment and diurnal flow. Alum dosage and peak flow rates were set based on previous plant operating experiences. The carbon contactors were run in a two-stage counter-current mode. The carbon contactor effluent was filtered through a tri-media filter.

Table 6-3 presents pertinent operating conditions and a tabulation of average plant influent, chemical treatment effluent and plant effluent quality for this operating period. Figure 6-4 shows the plant effluent history curves. Since the filter was not brought on stream until April 13, data for the final carbon contactor was plotted prior to this date. The substantial drop in suspended solids and total phosphorus after the filter was brought on stream is obvious.

Two carbon dosages were used during this operating period. From April 15 to May 5, the average carbon dosage was 294 mg/ℓ. Before and after this period the carbon dosage averaged 104 mg/ℓ. At the higher dosage, lower COD and SCOD values were achieved. COD and BOD_5 data for the two different dosage periods are broken out in Table 6-3 to show the effect of the higher carbon dosage.

During this operating period two diurnal flow variations were used, 2:1 and 3:1 peak to minimum flow. As seen in Figure 6-4, the 3:1 variation did significantly effect the plant effluent quality and consistency.

During the 2:1 peak:minimum flow period, the average plant flow was 35 gpm. The peak flow ranged from 40 to 44 gpm and the minimum flow was 20 to 24 gpm. During this period, the plant effluent averaged 4 mg/ℓ SS, 6.3 mg/ℓ COD and 2.7 mg/ℓ SCOD. During the 3:1 peak:minimum flow period, the average plant flow was 41 gpm. The peak flow ranged from 55 to 58 gpm and the minimum flow was about 20 gpm. During this period the plant effluent averaged 13 mg/ℓ SS, 11 mg/ℓ COD and 5 mg/ℓ SCOD. The slight difference in effluent SCOD

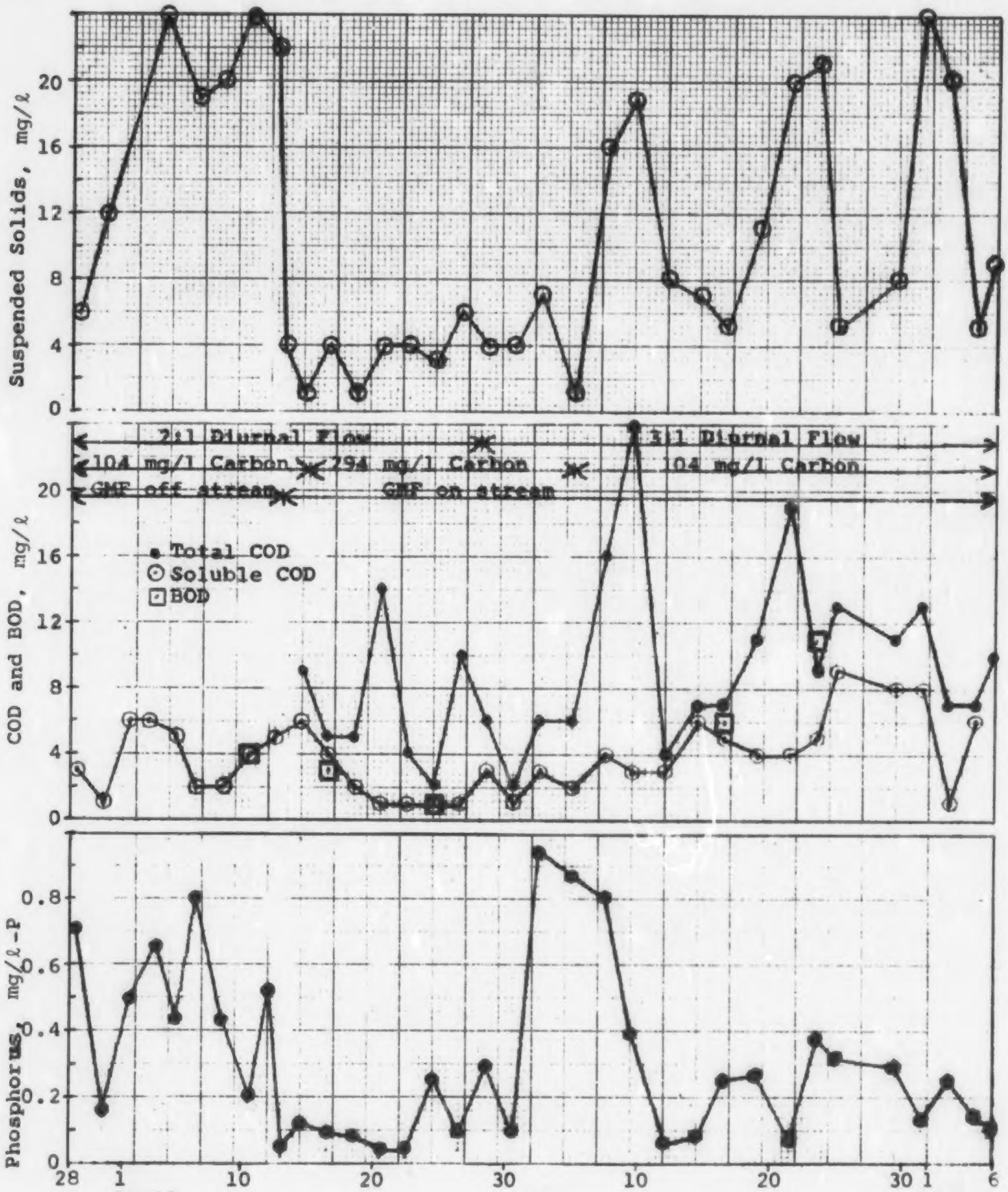
TABLE 6-3: OPERATING PERIOD III: AVERAGE OPERATING CONDITIONS TREATMENT RESULTS MARCH 28 TO JUNE 7, 1973

	Operating Conditions	
Chemicals:	Alum @ 14.0 mg/ℓ - Al	
Average Hydraulic Loading:	Chemical Contactor	0.38 gpm/sq ft
	First Stage Carbon	0.51 gpm/sq ft
	Second Stage Carbon	0.44 gpm/sq ft
	GM Filter	2.0 gpm/sq ft
	Peak to minimum	2:1 and 3:1
Carbon Contacting Mode:	Two-stage counter-current	
Carbon Dosage:		294 mg/ℓ and 104 mg/ℓ

Average Process Treatment Results	Raw Wastewater	Chemical Effluent	Plant Effluent
Turbidity, JTU	53	11	2
Suspended Solids, mg/ℓ	174	29	9
% Transmittance	-	-	97
Total P, mg/ℓ P	5.03	0.67	0.2
Soluble Total P, mg/ℓ P	3.87	0.14	0.1
Total σPO_4, mg/ℓ P	2.91	0.28	0.1
Soluble σPO_4, mg/ℓ P	2.52	0.08	0.07
COD, mg/ℓ	143	41	5/11[a]
SCOD, mg/ℓ	38	23	2.0/5.0[a]
BOD, mg/ℓ	112	26	3/8[a]
pH, units	7.1	7.3	7.5

a - 294 mg/ℓ of PAC/104 mg/ℓ of PAC

FIGURE 6-4: EFFLUENT HISTORY CURVE FOR OPERATING PERIOD III (MARCH 28 - JUNE 7, 1973)



was due to a substantially higher carbon dosage during the first part of this period. (See Figure 6-4.)

The substantial effect of flow variation noted above is felt to have been caused by one or two reasons. The first relates to the significantly higher peak hydraulic loading on the chemical treatment unit clarification zone. During the 3:1 flow variation period, the peak hydraulic loading was about 0.57 gpm/sq ft, whereas during the 2:1 flow variation period, the peak hydraulic loading was about 0.42 gpm/sq ft. The 40 percent higher peak hydraulic loading resulted in more chemical floc escaping the clarification zone. The second possible reason for deteriorated granular media filter performance may have been due to the actual diurnal flow variation experienced. Because from 4 to 5 gpm sample streams were taken from the overflows of the three preceding solids-contact units, the actual diurnal flow variation to the filter was well in excess of the 2:1 and 3:1 peak:minimum plant feed flow. The filter actually experienced 4:1 and 6:1 diurnal flow variations. The high variation may have contributed to the deteriorated performance.

From the plant effluent phosphorus history curve shown in Figure 6-4, it is seen that during the period, May 2 to 8, abnormally high values of effluent phosphorus were experienced. The reason for this abnormality in phosphorus removal was determined to be due to low alum dosages. Average daily alum dosages were found to be about 40 percent of the average dosage for Period III of 14 mg/ℓ aluminum. The average plant effluent phosphorus value shown in Table 6-3 does not include values for the period May 2 to May 8.

During Period III, overall plant SCOD removal was 90 percent. The alum treatment step removed 44 percent and the carbon step removed 56 percent of the total SCOD removed. Suspended solids were reduced by 95 percent, COD by 94 percent, and total phosphorus by 96 percent. In general, the plant effluent was of very good quality.

Operating Period IV

Operating Period IV was from June 10 to August 13, 1973. During this period $FeCl_3$ pretreatment and 2:1 peak:minimum diurnal flow pattern were used. The carbon contactors were operated in two-stage counter-current mode and the effluent was filtered through a tri-media filter bed.

Table 6-4 presents pertinent operating conditions and a tabulation of average influent, chemical treatment effluent, and plant effluent quality. Figure 6-5 is a plant effluent history curve for this period.

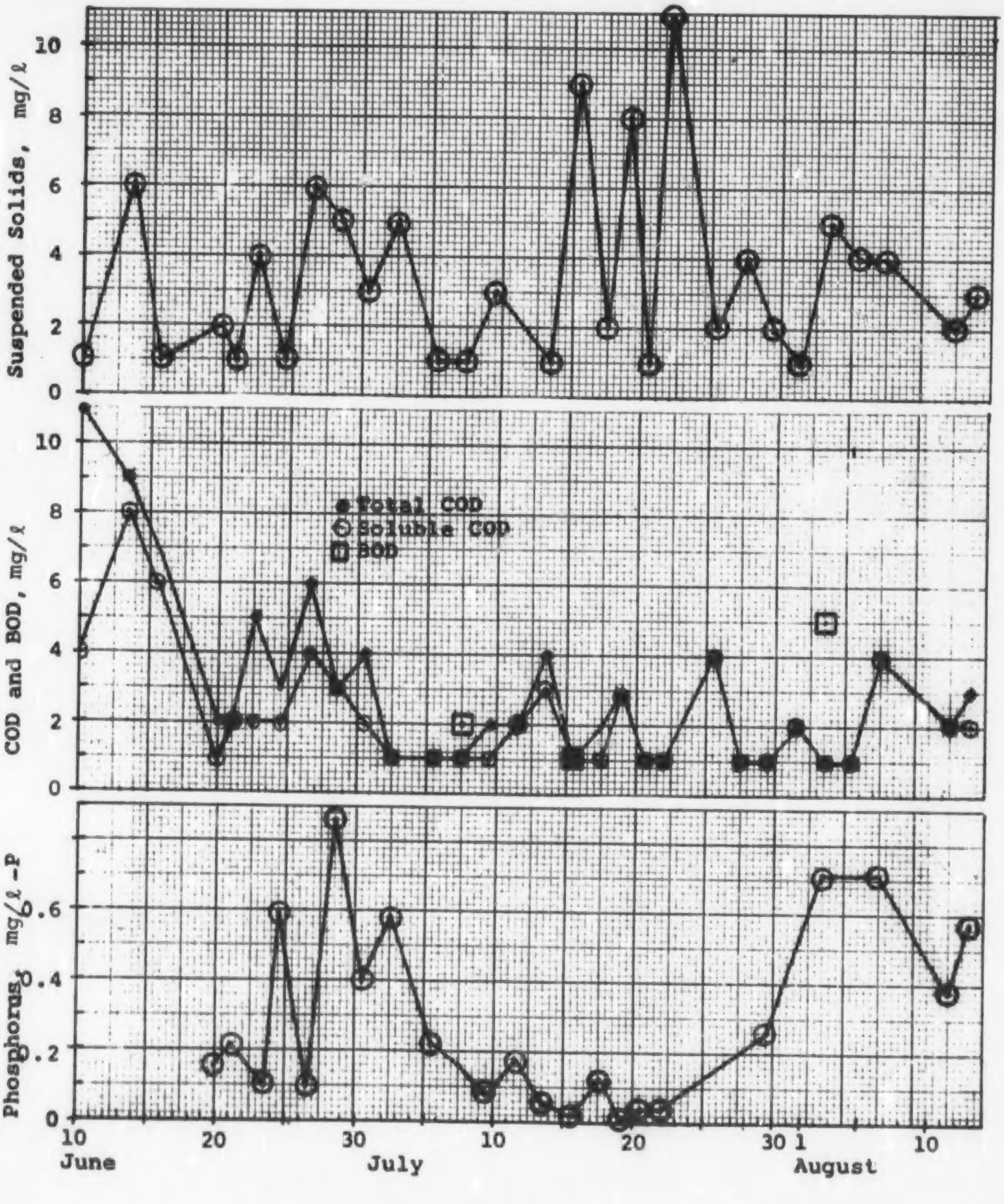
TABLE 6-4: OPERATING PERIOD IV: AVERAGE OPERATING CONDITIONS AND TREATMENT RESULTS JUNE 10 TO AUGUST 13, 1973

Operating Conditions

Chemical:	$FeCl_3$ @ 96 mg/ℓ	
Average Hydraulic Loading:	Chemical Contactor	0.38 gpm/sq ft
	First Stage Carbon	0.52 gpm/sq ft
	Second Stage Carbon	0.46 gpm/sq ft
	GM Filter	2.1 gpm/sq ft
	Peak to minimum	2:1
Carbon Contacting Mode:	Two-stage counter-current	
Carbon Dosage:		95 mg/ℓ

Average Process Treatment Results	Raw Wastewater	Chemical Effluent	Plant Effluent
Turbidity	49	42	3
Suspended Solids, mg/ℓ	118	26	3
% Transmittance	-	-	99
Total P, mg/ℓ	4.93	0.52	0.29
Soluble Total P, mg/ℓ P	3.80	0.11	0.20
Total σPO_4, mg/ℓ P	3.76	0.38	0.19
Soluble σPO_4, mg/ℓ P	3.35	0.05	0.13
COD, mg/ℓ	128	28	3
SCOD, mg/ℓ	35	19	2
BOD, mg/ℓ	97	7	3
pH, units	7.2	7.3	7.6

FIGURE 6-5: EFFLUENT HISTORY CURVE FOR OPERATING PERIOD IV (JUNE 10 - AUGUST 13, 1973)



The two periods of abnormally high plant effluent phosphorus were determined to be due to low $FeCl_3$ dosages.

As seen in Figure 6-5, there was about a 10 day period before plant effluent SCOD reached the low value of about 2 mg/ℓ. Obviously the chemical treatment unit was not removing substantial SCOD until after about 10 days, or by June 19. This fact is shown in Figure 6-6, which indicates that no removal of SCOD was experienced for the first 5 days and then increasing SCOD removal, up to 50 percent, by the 12th day. As will be discussed shortly, anaerobic biological activity existed in the chemical treatment unit. It is possible that anaerobic biological activity also removed some SCOD in the chemical treatment unit. This would explain why it took almost 2 weeks before substantial SCOD removals were observed after start-up of the unit for the $FeCl_3$ campaign.

During previous contract work, a "post precipitation" problem was experienced when $FeCl_3$ was used as the coagulant for raw wastewater in a solids-contact treatment unit.[1] The problem was conjectured to have been due to reduction of ferric to ferrous iron in the chemical treatment unit and subsequent oxidation back to ferric iron in composite samples and plant effluent clearwell. The cause of the reduction of iron was suggested to be due to anaerobic conditions existing in the chemical treatment unit. This same problem reappeared during the current study, but with an added negative effect.

Chemical treatment unit average composite sample effluent turbidity shown in Table 6-4 for Period IV was 42 JTU's. During this same period, the average of at least six grab sample turbidities each day was about 9 JTU's. This difference is similar to that reported for previous studies.[1]

TABLE 6-5: PLANT GRAB-SAMPLE SOLUBLE IRON PROFILES

Date	August 8	August 28
Process Stream:		
Raw Wastewater	0	0
Chemical Treatment Effluent	11	7.0
First Stage Carbon Effluent	--	5.5
Second Stage Carbon Effluent	--	2.5
Plant (GMF) Effluent	0	0

FIGURE 6-6: SOLUBLE CHEMICAL OXYGEN DEMAND REMOVAL PATTERN DURING INITIAL START-UP OF CHEMICAL CONTACTOR WITH $FeCl_3$

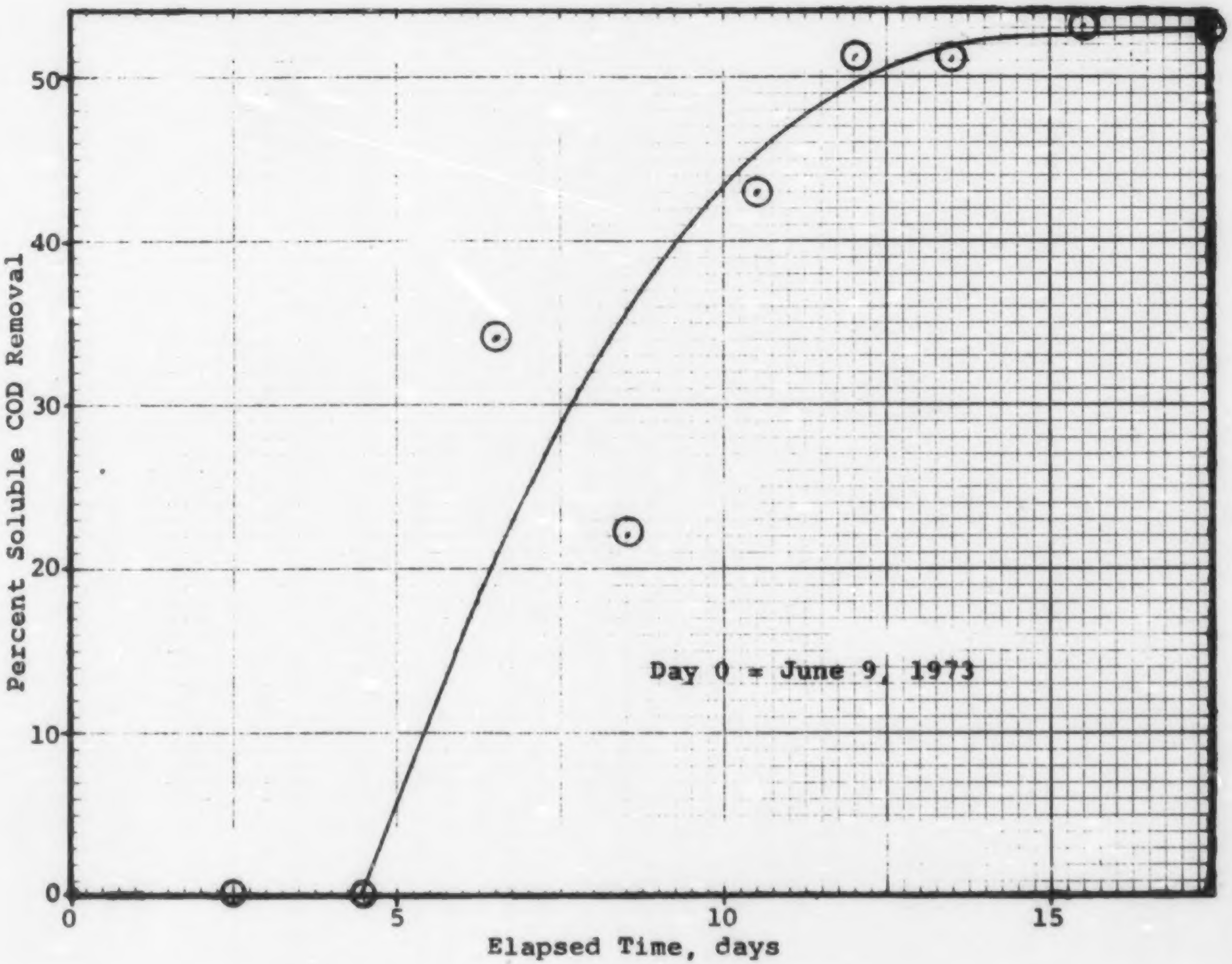


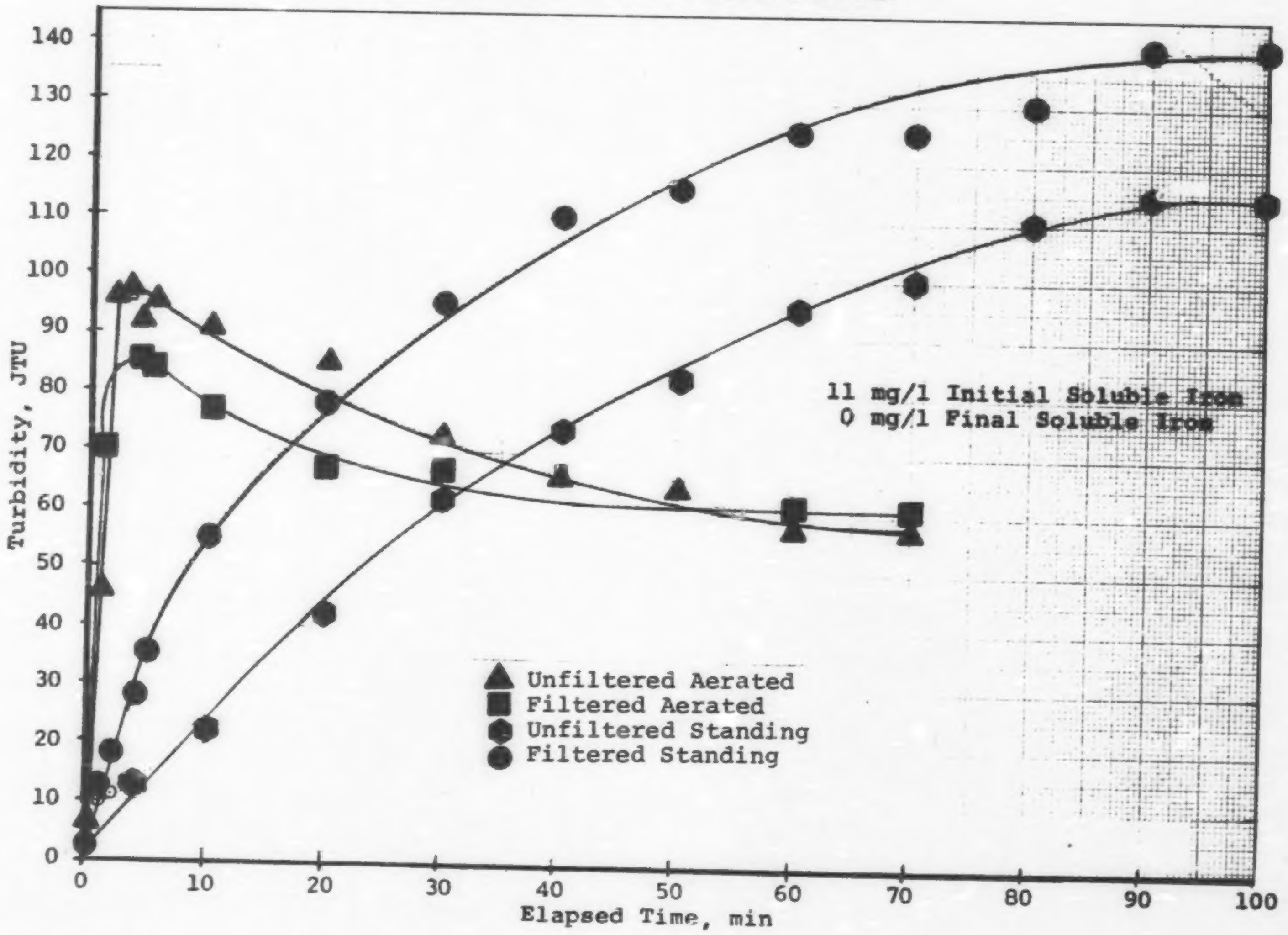
Table [illegible]6 presents two plant soluble iron profiles. These show high concentrations of soluble iron in the chemical treatment unit effluent. It was subsequently removed in the carbon and granular media filter steps. The only possible external source of soluble iron would have been in the $FeCl_3$ feed stock solution. This was considered to be unlikely since high purity $FeCl_3$ powder was used (95 percent purity). To verify this belief, soluble iron determinations were made of the flash-mix effluent which contained the $FeCl_3$ feed. No soluble iron was found. Thus it was concluded that all iron reporting to the chemical treatment unit was insoluble ferric precipitate (floc). Since the effluent pH was about 7.0 to 7.5, it was assumed that any soluble iron present was in the ferrous form. This assumption was substantiated by the following bench scale test. A grab sample of chemical treatment unit effluent was split into two aliquots. One of the aliquots was filtered through a 0.45 micron membrane filter in the absence of air. The filtered and unfiltered samples were split into two aliquots and one of each aerated vigorously and one of each left sitting quiescently. Turbidities of the four samples were determined versus elapsed time and were plotted in Figure 6-7. It is readily apparent that gross turbidity increases were observed for all samples. Vigorous aeration resulted in near instantaneous precipitation of a turbidity exerting material. Soluble iron determinations after turbidity equalization showed none of the initially present 11 mg/ℓ of soluble iron. It was therefore assumed that soluble iron was originally present in the ferrous form.

The pertinent implication of the above assumption and previously mentioned observations is that iron was reduced from the ferric to ferrous form within the chemical treatment unit.

An additional problem encountered during this period was the appearance of a black, finely divided precipitate in the chemical clarifier underflow sludge. This precipitate was not detected in the clarifier effluent. As will be discussed in Section VII, the presence of this black precipitate gave the sludge a septic appearance. However, no signs of gross septicity, such as "rotten egg" odors or gaseousness, were found. Gravity thickener performance was not adversely affected, but the finely divided precipitate created a difficult problem with the vacuum filtration of ferric-primary sludge. Upon acidification of samples of filter cake gross sulfide odors (rotten egg) were observed.

Several grab sample profiles of all process streams, thickener overflow and vacuum filter feed and filtrate all

FIGURE 6-7: EFFECT OF AERATION ON IRON PRECIPITATION



indicated nominal to non-existent amounts of soluble sulfides. The only plausible explanation of where the copious amount of sulfide could have been produced was that it occurred in the chemical treatment unit and thickener. This implies the presence of anaerobic biological activity in general and of anaerobic sulfur reducing bacterial activity specifically.

In summary, it is concluded that anaerobic biological activity in the chemical treatment unit resulted in the reduction of iron and sulfur with the resultant precipitation of FeS and leakage of soluble iron in the effluent. As previously noted some biological removal of SCOD was also probably experienced.

With $FeCl_3$ treatment and two-stage counter-current carbon contacting, the pilot plant removed 94 percent SCOD. The chemical treatment step removed 48 percent and the carbon step 52 percent of the total SCOD removed. A 97 percent reduction in total COD was achieved. A 94 percent reduction of total phosphorus was achieved.

Granular media filtration was very effective. Note that the total COD values shown in Figure 6-5 are almost always equal to SCOD values.

During operating Period IV, the pilot plant consistently produced excellent quality effluent.

Operating Period V

Operating Period V was from August 14 to September 15, 1973. The major treatment variation during this period was the use of a reduced $FeCl_3$ coagulant dosage and a polyelectrolyte floc strengthener.

During Period IV, the ratio of $FeCl_3$ fed to phosphorus was 3.7 moles per mole. Though as much as 1/3 of the ferric iron may have been solubilized, reducing this ratio to 2.5, the 0.29 mg/ℓ total phosphorus in the effluent was considered quite low. Thus during Period V it was decided to significantly reduce the $FeCl_3$ dosage.

Previous experience had indicated that 0.25 mg/ℓ of Altasep 2A2* was an optimum dosage for 50 to 75 mg/ℓ $FeCl_3$ dosages.[1]

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*A product of Atlas Chemical Industries Incorporated, Wilmington, Delaware.

Using this polyelectrolyte at about \$1.50/lb, a reduction of 7.5 mg/ℓ $FeCl_3$ at 5¢/lb would be required for a cost trade off. For Period V a target of about 70 mg/ℓ $FeCl_3$ was set and as seen in Table 6-6 the average dosage was 62 mg/ℓ.

The carbon contactors were operated in the two-stage countercurrent mode. From August 24 on, regenerated carbon was used.

Table 6-6 presents pertinent operating conditions and average influent, chemical treatment effluent and plant effluent quality. Figure 6-8 is a plant effluent history curve. As was experienced during the previous operating Period (IV) very consistent treatment was achieved. Comparison of average plant effluent qualities during Periods IV and V (Tables 6-4 and 6-6) shows that the only significant change was an increase of phosphorus from 0.29 to 0.67 mg/ℓ. Comparison of chemical treatment unit and plant effluent phosphorus data shown in Table 6-6 indicates a slight reduction in total phosphorus across the two-stages of carbon contactors and the granular media filter. However, a substantial increase in soluble phosphorus occurred. It can only be presumed that the anaerobic conditions in the carbon contactors resulted in additional reduction of Fe^{+3} resulting in release of phosphorus. The same phenomenon was experienced during Period IV (Table 6-4).

The iron post-precipitation problem identified during Period IV persisted during Period V. As noted previously (Table 6-5) soluble iron was removed across the carbon contacting stages and the filter. An equilibrium adsorption isotherm test indicated that substantial soluble iron was adsorbed on virgin carbon.

On August 24, reuse of thermally regenerated carbon was started. As seen in Figure 6-8, a slight increase in plant effluent COD was experienced when compared to the virgin carbon period.

During Period V the pilot plant removed 90 percent of the SCOD. The chemical treatment step removed 52 percent and the carbon step removed 48 percent of the total SCOD removed. Suspended solids removal was 90 percent and 98 percent removal of total COD was experienced. Only 86 percent removal of total phosphorus was achieved because of a reduced $FeCl_3$ dosage. Again the granular media filter was very effective in producing low levels of suspended solids in the effluent. The final effluent was of excellent quality and the operation very consistent for this period.

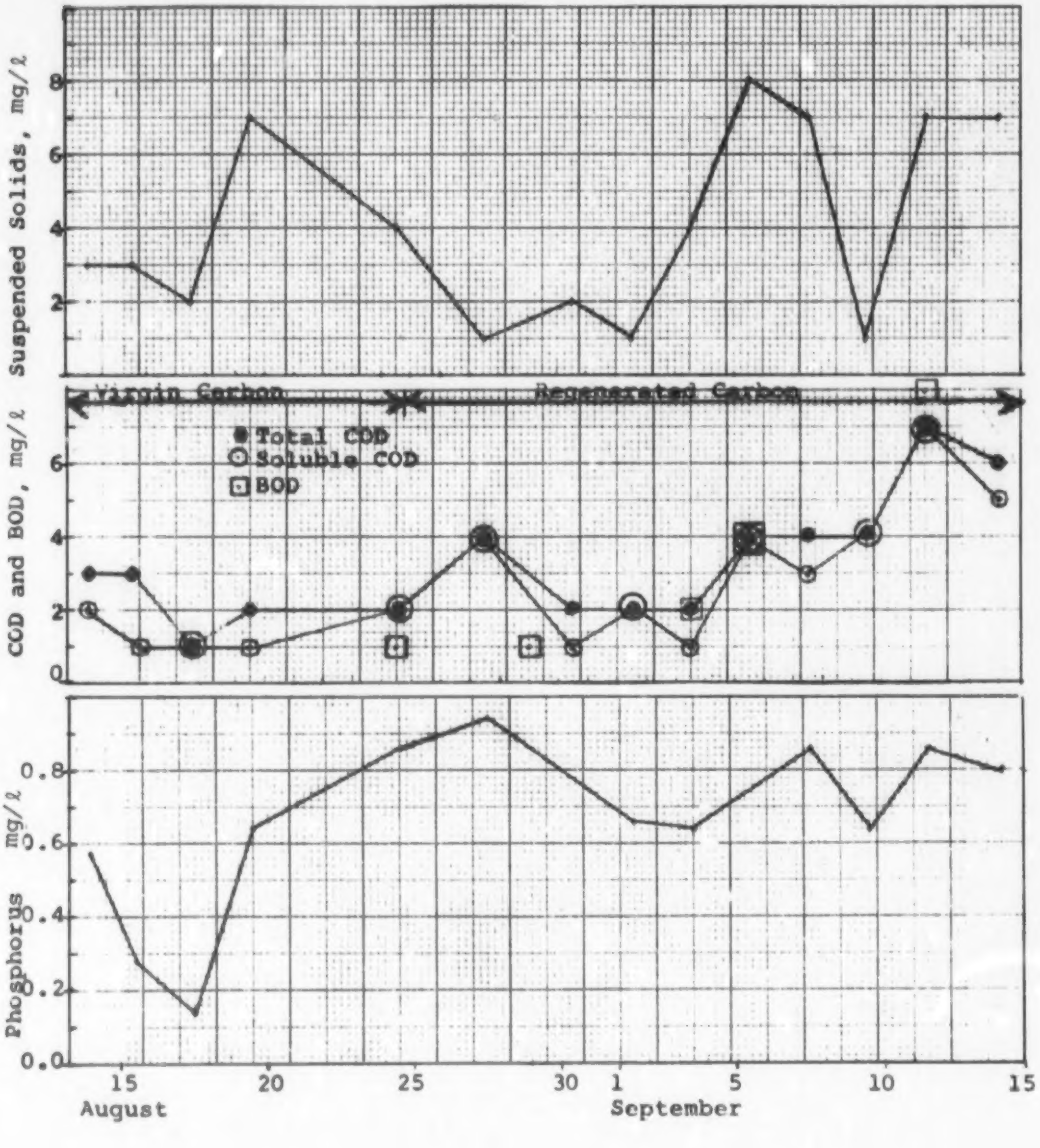
TABLE 6-6: OPERATING PERIOD V: AVERAGE OPERATING CONDITIONS AND TREATMENT RESULTS AUGUST 14 TO SEPTEMBER 15, 1973

Operating Conditions

Chemicals: FeCl$_3$:	@ 62 mg/ℓ	
Polymer:	@ 0.25 mg/ℓ	
Average Hydraulic Loading:	Chemical Contactor	0.39 gpm/sq ft
	First Stage Carbon	0.52 gpm/sq ft
	Second Stage Carbon	0.45 gpm/sq ft
	GM Filter	2.7 gpm/sq ft
	Peak to minimum	2:1
Carbon Contacting Mode:	Two-stage counter-current	
Carbon Dosage:		98 mg/ℓ

Average Process Treatment Results	Raw Wastewater	Chemical Effluent	Plant Effluent
Turbidity, JTU	43	35	2
Suspended Solids, mg/ℓ	166	35	4
% Transmittance	-	-	99
Total P, mg/ℓ P	4.94	0.80	0.67
Soluble Total P, mg/ℓ P	3.67	0.24	0.61
Total σPO_4, mg/ℓ P	4.31	0.51	0.63
Soluble σPO_4, mg/ℓ P	3.76	0.10	0.59
COD, mg/ℓ	133	28	3
SCOD, mg/ℓ	30	16	3
BOD, mg/ℓ	58	16	3
pH, units	7.2	7.2	7.3

FIGURE 6-8: EFFLUENT HISTORY CURVE FOR OPERATING PERIOD V (AUGUST 14 - SEPTEMBER 15, 1973)



Operating Period VI

On September 15, the pilot plant was taken off stream to terminate the $FeCl_3$ chemical treatment campaign. Plant modifications were implemented to allow for returning to lime chemical treatment. As will be developed in later Sections, additional lime sludge thickening experience was considered desirable. In addition, plant SCOD removal data for lime chemical treatment ahead of two-stage counter-current carbon treatment was desired. On September 18, the chemical treatment unit was started up with lime and on September 20, acid (H_2SO_4) neutralization was started. Instrumentation, equipment, and operational problems were numerous. The seven months idle time of some instruments, controls and equipment resulted in extensive maintenance, calibration and repair work. Coupled with this was the fact that the plant operating staff at the time had no experience with lime treatment. This was due to reassignment of two key senior technicians versed in plant operations and normal turn over of part time and full time operation personnel. Consequently, retraining of the new operating personnel was required. After about one week (on September 27), the decision was made to abort the lime chemical treatment study. Time was of the essence and the major priority at that time was to regenerate and evaluate the reuse of carbon. Erratic operation of the chemical treatment unit using lime could not be tolerated, even for a few weeks.

The last operation, Period VI, was from September 28 to November 8, 1973. Alum pretreatment with polymer addition was used with a 2:1 diurnal flow pattern. The carbon contactors were again operated in the two-stage counter-current mode.

Table 6-7 presents pertinent operating conditions and average plant influent, chemical treatment effluent and plant effluent quality. Figure 6-9 is the plant effluent history curve for this operating period. Since emphasis was placed on carbon regeneration which required considerable laboratory time, no phosphorus analyses were done. Operational and sampling problems were experienced from October 9 to October 15, and thus, no laboratory analyses were made.

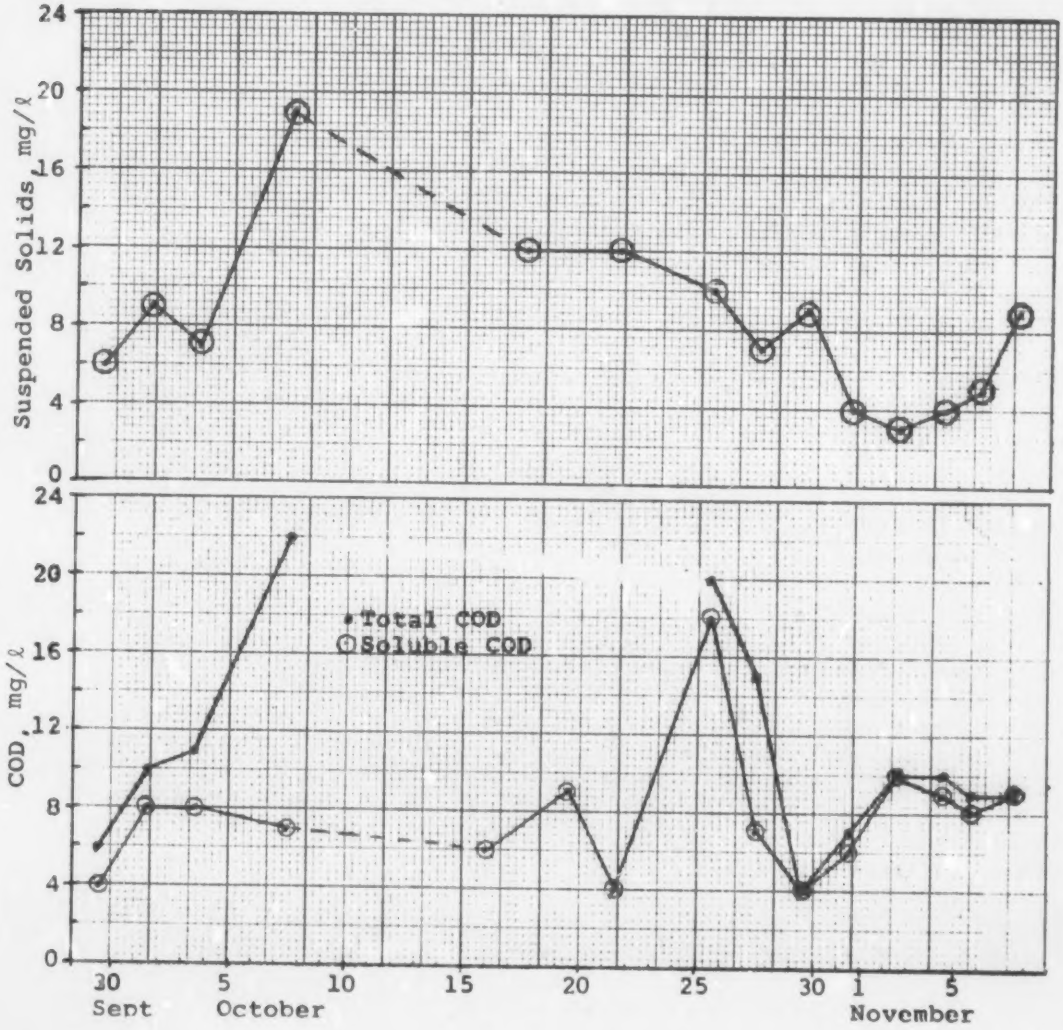
Since low carbon dosages were used, residual COD was slightly higher than for previous periods. However, 82 percent removal of SCOD was achieved. The chemical contactor removed 56 percent of the total SCOD removed while 44 percent was removed in the carbon system. Suspended solids and total COD removals were 97 and 94 percent, respectively.

TABLE 6-7: OPERATING PERIOD VI: AVERAGE OPERATING CONDITIONS AND TREATMENT RESULTS SEPTEMBER 28 TO NOVEMBER 8, 1973

	Operating Conditions	
Chemicals: Alum:	@ 9.6 mg/ℓ - Al	
Polymer:	@ 0.25 mg/ℓ	
Average Hydraulic Loading:	Chemical Contactor	0.49 gpm/sq ft
	First Stage Carbon	0.63 gpm/sq ft
	Second Stage Carbon	0.55 gpm/sq ft
	GM Filter	3.5 gpm/sq ft
	Peak to minimum	2:1 and 3:1
Carbon Contacting Mode:	Two-stage counter-current	
Carbon Dosage:		62 mg/ℓ Fixed Carbon

Average Process Treatment Results	Raw Wastewater	Chemical Effluent	Plant Effluent
Turbidity, JTU	59	8	8
Suspended Solids, mg/ℓ	263	26	8
% Transmittance	-	-	97
COD, mg/ℓ	179	49	11
SCOD, mg/ℓ	44	24	8
BOD, mg/ℓ	187	37	6
pH, units	7.1	7.3	-

FIGURE 6-9: EFFLUENT HISTORY CURVE FOR OPERATING PERIOD VI
(SEPTEMBER 28 - NOVEMBER 8, 1973)



SOLIDS-CONTACT TREATMENT UNITS PERFORMANCE

Solids-contact treatment units were employed for two of the three liquid wastewater treatment unit operations used in the PAC-PCT system studied. Unit processes of chemical precipitation, coagulation, flocculation, solids-contacting, adsorption, biological organic removal, and gravity sedimentation were accomplished in the two unit operations broadly identified as chemical treatment-clarification and carbon contacting-clarification. To say that the successful, or unsuccessful, performance of the PAC-PCT pilot plant depended to a large degree on the performance of the solids-contact units would be an understatement.

The purpose of this sub-section is to evaluate two aspects of the performance of the solids-contact units used. Operational stability and clarification effectiveness for the three general types of applications evaluated will be discussed. These three applications were lime treatment, alum or $FeCl_3$ chemical treatment of a raw municipal wastewater and carbon treatment of the chemically treated effluent.

The need to emphasize this topic is based on a few negative pilot plant experiences and a few published negative judgement comments generally based on very indirect experience.[4],[5],[6],[7] On two occasions known to the authors, negative pilot plant experiences were the result, in part, of using inappropriate pilot plant units offered by an equipment manufacturer. In another instance with a prototype unit, an inadequately designed sludge underflow pump caused a major problem.

On the positive side, it should be noted that hundreds of solids-contact treatment units, of various manufacturers, have been successfully applied to municipal and industrial wastewater applications. Typical industrial wastewater applications are for treatment of metalurgical wastes (e.g., metal finishing, fluoride removal, rinse water treatment), steel mill wastes and pulp and paper wastes (e.g., white water, color removal).

Most of these applications are not at constant flow but rather for stepped-variable flow.

The major concern with solids-contact treatment in municipal wastewater applications is the presence of a large inventory of sludge in the unit. Fear of excessively high shock hydraulic loadings or gross thermal upsets are common apprehensions. A real problem of iron and sulphur reduction was experienced in this study. This problem resulted from the detention and mixing of a large inventory of primary sludge at near neutral pH when ferric chloride was used. No operational or performance problems were experienced when using alum treatment at near neutral pH, or lime treatment at a pH of 10.0 to 11.2.

To enable evaluation of the effect of flow, process and operational variations on the clarification performance, and operational stability of the specific solids-contact units used in this study (see Figure 4-3), approximate sludge pool depths and concentrations were monitored six times each day. Sludge pool depth was approximated by observing and recording the highest sample tap at which sludge appeared. Sample taps were located at about 13, 31, and 58 inches up the tank wall from the tank bottom. These taps were designated as the bottom, middle, and top.

Sludge concentration was routinely determined by an indirect 5 or 10 minute settling test. The procedure used was as follows: A 2 liter graduated cylinder was filled to 2.0 liters of sludge. The sludge was allowed to quiescently settle for 5 to 10 minutes. The volume of settled sludge was observed and the percent volume it represented noted. These values were reported as the 5 (or 10) minute percent settled sludge volume. The time was chosen such that the subsidence rate of the sludge-liquid interface at that time was very small (e.g. less than about 0.2 ℓ/min). For alum or ferric-primary sludges, 10 minutes was determined to be required, whereas for lime-primary and carbon sludges 5 minutes was sufficient.

Suspended solids concentrations were found to be directly related to percent settled sludge volumes. For alum or ferric-primary sludges the relationship was 0.05-0.15 g/ℓ per percent settled sludge volume, with an average of 0.10. An average of 0.15 g/ℓ per percent settled sludge volume was found for carbon sludges with a range of from 0.09 to 0.22. For lime-primary sludge the average was 0.19 g/ℓ per percent settled sludge volume with a range of from 0.06 to 0.47.

The 5 or 10 minute settling test also provided an approximate and normally conservative estimate of the sludge interface subsidence rate. The percent settled sludge volumes reported were directly related to a height the sludge interface fell in the given time period. The 2.0 liter mark on graduate cylinders used was about 1.3 ft or 16 inches. Thus a 50 percent settled sludge volume meant that the sludge interface had settled 16-(16 x 50/100) or 8 inches during the test period. A 25 percent settled sludge volume meant the sludge interface had settled 16-(16 x 25/100) or 12 inches during the test period. The general expressions for sludge interface subsidence rates, in inches/minutes, for the two time periods used was:

5-minute test -- 3.2 (1 - %Vol/100)

10-minute test -- 1.6 (1 - %Vol/100)

Recall that 1.6 inches/minutes equals 1.0 gpm/sq ft. The operational utility of this approximation of sludge settling rate can be appreciated by the following example: For a known clarification zone geometry, the average rise rate at any sample depth is known for a given flow to the clarifier. For illustrative purposes, assume that in the carbon contactor the peak daily rise rate at the middle (31 inches deep) sample tap was 1 gpm/sq ft (1.6 inches/minute). The maximum desired percent settled sludge volume for a 5 minute test at this sample tap would have been

$$1.6 = 3.2\left(1 - \frac{\%\ \text{Vol}}{100}\right)$$

$$\text{or} \quad \%\ \text{Vol} = 50$$

If the sludge pool concentration was in excess of 50 percent volume, the slurry pool would have been expected to expand (increase its depth) to a point where about 50 percent volume existed.

Lime-Primary Sludge Application

The chemical contactor-clarifier was operated under two flow patterns during lime treatment studies. First was constant flow at about 1.0 gpm/sq ft and the second at 2:1 peak: minimum diurnal flow with a peak of about 1.3 gpm/sq ft and an average of about 1.0 gpm/sq ft. Figures 6-10 and 6-11 show the daily high and low sludge levels (sample taps) and the effluent suspended solids values for the two flow patterns. Comparison of effluent suspended solids concentration ranges and averages indicated little difference in the range of suspended solids, but do indicate slightly lower suspended solids values during the constant flow period. The daily variation in sludge pool depth was consistently greater for the 2:1 diurnal flow period (Figure 6-11). The sludge pool interface rose and fell at least 2-1/2 ft and possibly as much as 3 to 4 ft. This resulted in a 3 to 4 ft sludge blanket existing at the peak flow of about 1.3 gpm/sq ft. At low flow, about 0.60 gpm/sq ft, no sludge blanket existed. Recall that a sludge level at the

FIGURE 6-10: DAILY VARIATION IN LIME PRIMARY SLUDGE LEVEL AND EFFLUENT SUSPENDED SOLIDS AT CONSTANT FLOW

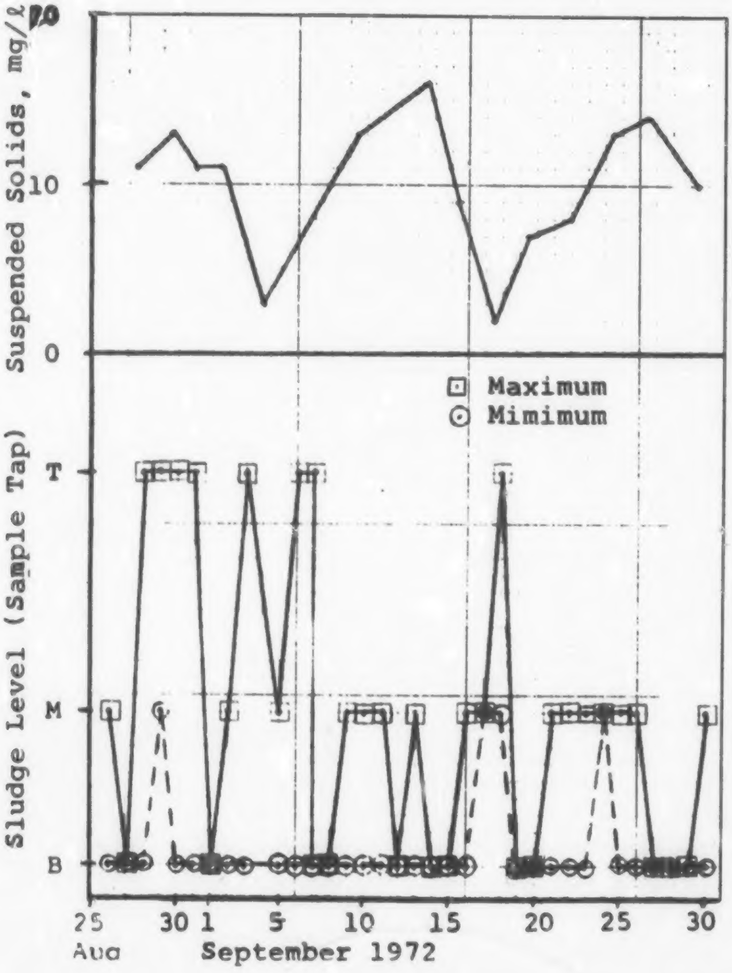
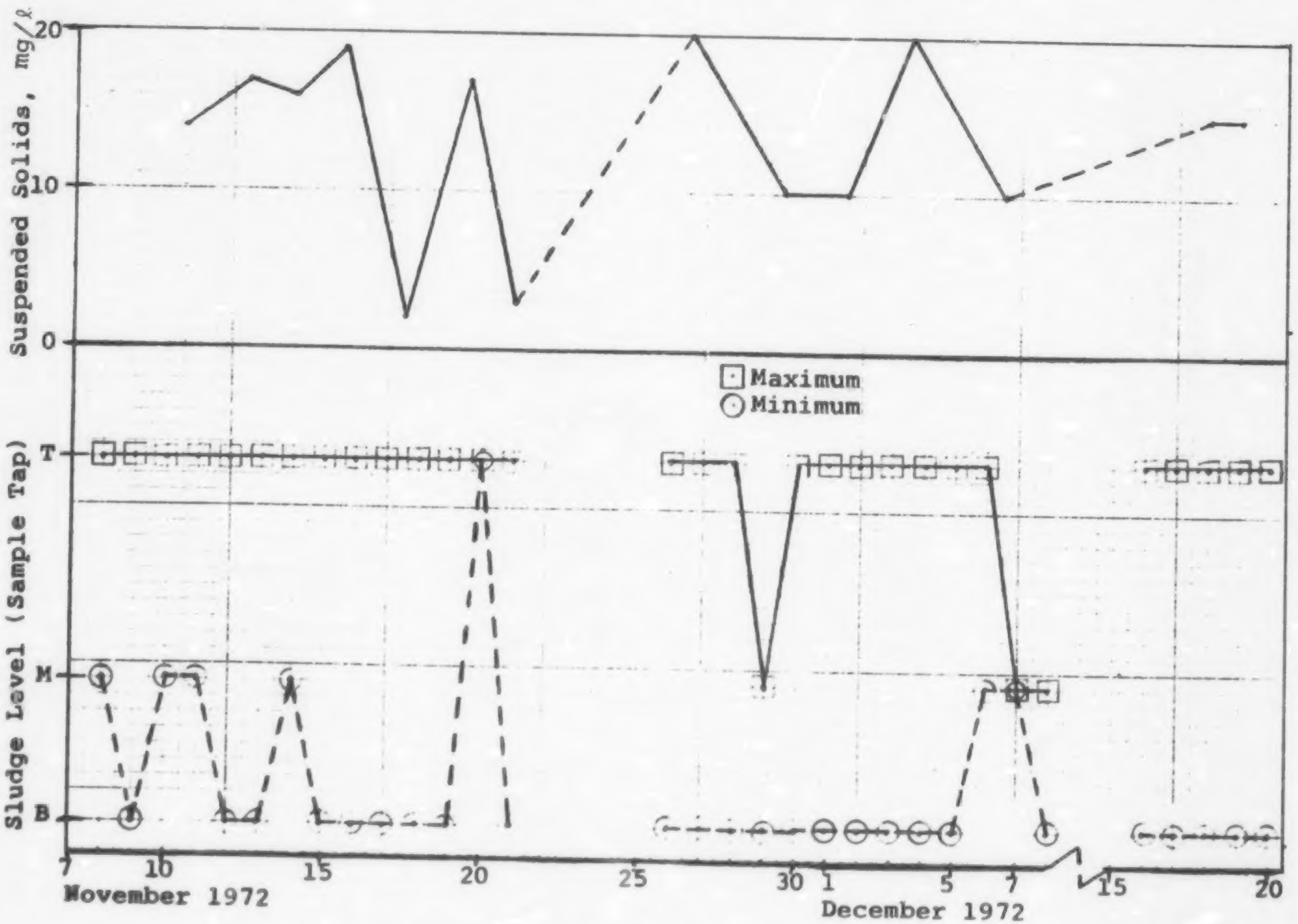


FIGURE 6-11: DAILY VARIATION IN LIME PRIMARY SLUDGE LEVEL AND EFFLUENT SUSPENDED SOLIDS AT 2:1 DIURNAL FLOW



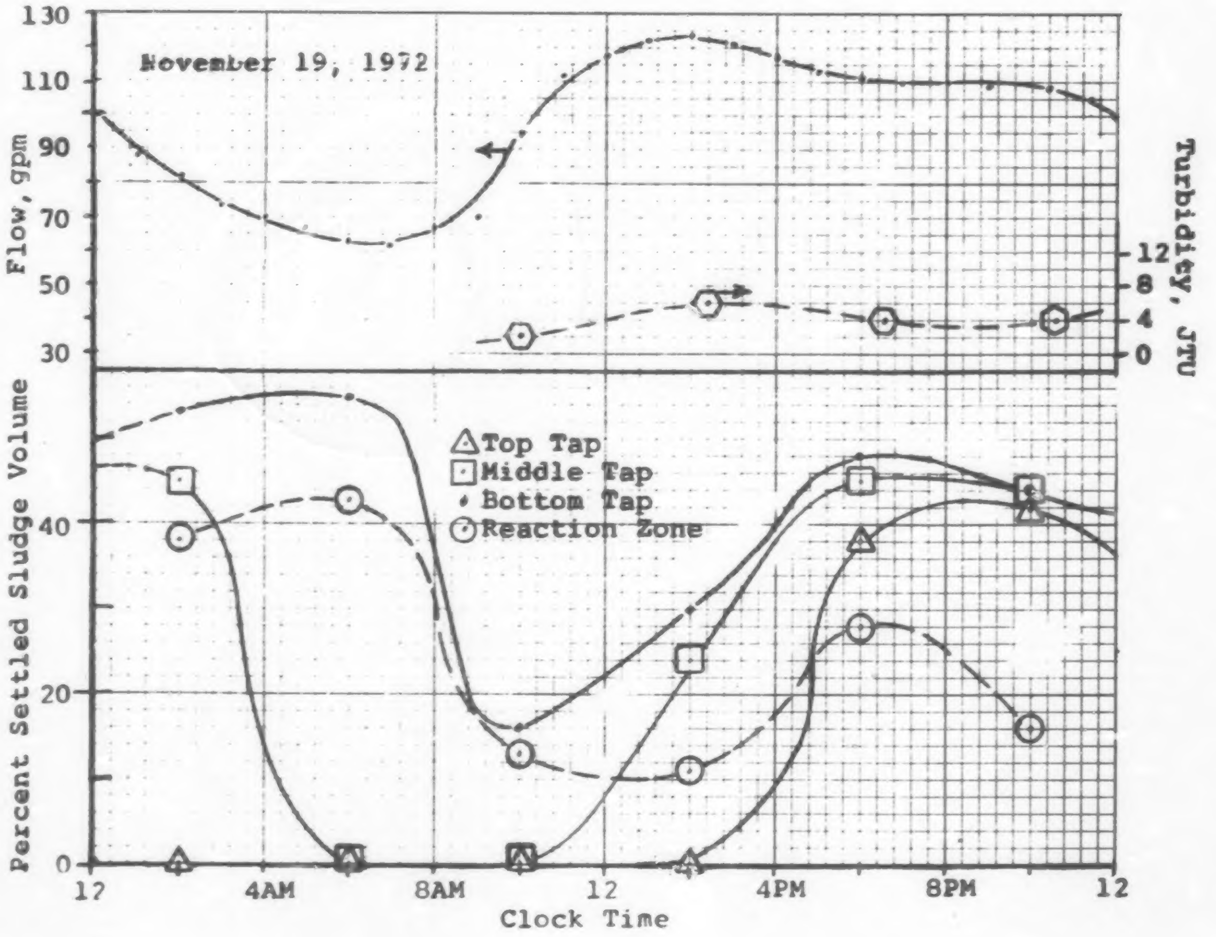
middle tap was required for accomplishing solids-contacting without sludge-blanket clarification. Sludge-blanket operation did not exist until the sludge pool interface rose about to the top tap.

During the constant flow period, shown in Figure 6-10, a sludge-blanket existed for very short periods on just a few days. The daily variations in sludge pool levels during this constant flow period, with an average hydraulic loading of about 1.0 gpm/sq ft, were obviously not due to flow variations. Rather, they represent operational variations, mainly in sludge blowdown frequencies (rates). At constant flow (1.0 gpm/sq ft), there were several week durations when daily sludge pool depth variations were as great as those shown in Figure 6-11 for a 2:1 diurnal flow pattern. Such a period occurred during most of January 1973, when failure of diurnal flow control instrumentation necessitated constant flow operation. During this period poor operation of the solids-contact unit resulted in average daily effluent suspended solids ranging from 5-80 mg/ℓ. The January 1973, monthly average was a high 31 mg/ℓ. One probable cause of operational difficulties during January 1973, was the colder than normal ambient (-5 to -10°F) and wastewater (11 to 13°C) temperatures.

On any given day, if the hydraulic loading, lime dosage, pumping turbine and sludge rake speeds were all constant and the sludge blowdown rate was at a constant value such that sludge withdrawal equalled solids production, then the concentration of sludge at any sample point in the solids-contact unit should have been constant. During this study, there were few days when all the above operational variables were constant. When they all were constant, very stable sludge level and constant suspended solids concentrations at the various sample taps were observed.

Figure 6-12 shows sludge concentration variations during a given day when lime dosage, sludge blowdown frequency and turbine and rake speeds were constant. The only imposed variable was diurnal flow. It is readily apparent that a consistent and somewhat predictable pattern of sludge level and concentration variations resulted from the diurnal flow effect. As flow was reduced to the minimum the sludge level dropped below the middle sample tap (31 inches) and approached the bottom tap (13 inches), and reaction zone taps. As the flow approached the daily peak the sludge pool expanded and the interface gradually rose over a 6 hour period. At about 6:00 P.M., the sludge pool interface was over 5-ft from the tank bottom and a fairly uniform concentration existed. The sludge-blanket was stable as indicated by the very low effluent turbidity.

FIGURE 6-12: TYPICAL HOURLY VARIATIONS IN LIME PRIMARY SLUDGE CONCENTRATION AND EFFLUENT TURBIDITY



The 6:00 P.M., test data for the top tap sample indicated about 38 percent settled sludge volume. For the 5 minute test used, this indicated the sludge pool interface settling rate would be approximately:

$$3.2\ (1 - 38/100) = 1.98 \text{ inch/min or } 1.24 \text{ gpm/sq ft}$$

The actual rise rate at the top tap location was 1.25 gpm/sq ft at 6:00 P.M. From the known geometry of the solids-contact unit and the 5 minute percent settled sludge volume data in Figure 6-12 at 6:00 P.M., it could be predicted that had the flow rate been increased to about 150 gpm and sustained there, the sludge-blanket would have expanded to the overflow launder and thus into the effluent. Had the operator expected such a high flow, he would have immediately reduced the turbine speed to minimize hydraulic turbulence and started manual blowdown of sludge to reduce the sludge inventory in the unit.

Obviously, the critical hydraulic design condition for this application relates to the sludge settling characteristics at the sustained high flow level expected. Previous contract studies indicated that at a constant loading of 1.3 gpm/sq ft stable operation and good suspended solids removal was achieved.[1] The maximum possible hydraulic loading for this application was not determined. It is the authors' judgement that a peak 2 hour hydraulic loading of about 1.5 gpm/sq ft would be possible.

Evaluation of daily results such as shown in Figure 6-12 indicated that during constant plant flow period (@ about 1.0 gpm/sq ft) the sludge pool concentration was very uniform. The difference between maximum and minimum 5 minute percent settled sludge volumes varied no more than 2 to 10 percent daily. During November and December 1972, when the plant was subjected to a 2:1 diurnal flow pattern, as shown in Figure 6-12, the difference between maximum and minimum 5 minute percent settled sludge volumes varied from 20 to 50 percent and averaged about 30 percent. On those days when lime dosage, turbine or rake speeds or blowdown frequencies were changed substantially, larger variations in the difference between maximum and minimum values were experienced.

Alum or $FeCl_3$-Primary Application

The treatment of Salt Lake City raw municipal wastewater with alum or $FeCl_3$ resulted in different operating characteristics than lime treatment. Based on previous contract work (reference 1) and other relevant field experience by one of the authors, the solids-contact unit was modified, as indicated in Figure 4-3, to include a sludge baffle plate. This baffle plate was for the purpose of separating the reaction zone and sludge concentrating zone of the solids contact unit. The objective of the baffle plate was to increase underflow solids concentration by reducing hydraulic turbulence in the sludge withdrawal zone. Another variation from previous contract work was to externally rapid-mix the alum or $FeCl_3$ just prior to the solids-contact unit. Thus, the main functions of the solids-contact unit in the current study was to flocculate via solids-contacting and clarify via gravity sedimentation.

Figure 6-13 shows the daily variation in sludge pool depth (sample tap location) and effluent composite suspended solids concentration for pilot plant operating Period III. The average alum dosage during this period was 14 mg/ℓ as aluminum. The 2:1 and 3:1 peak:minimum diurnal flow patterns used are shown in Figure 6-14. The effective clarification area of the solids-contact unit used was about 100 sq ft.

The data in Figure 6-13 shows that fairly consistent performance of the solids-contact unit was experienced. It is, however, obvious that clarification performance deteriorated substantially during the 3:1 diurnal flow period. Both the range and overall average effluent suspended solids concentration about doubled when comparing the 3:1 and 2:1 diurnal flow operation. As opposed to the other two periods shown in Figure 6-13, during the 3:1 diurnal flow period (without polyelectrolyte), there was almost always a sludge-blanket of at least 1-2 ft, even at minimum daily flows (about 0.22 gpm/sq ft).

Operational stability was considered poorer for the 3:1 diurnal flow period. During this 28 day period, the chemical treatment unit effluent was diverted from the carbon contactor due to unacceptably high suspended solids on five different occasions for from 2 to 7.5 hour duration. The average down time was just under four hours. No down time was required during the other two periods.

A clearer picture of the effect of diurnal flow variations can be seen by considering the fairly typical daily variations of sludge pool concentration and effluent turbidity shown in Figure 6-15. The sludge pool solids concentration is seen to be fairly uniform at all clarification zone sample locations for all three conditions shown.

On May 17, the sludge inventory was at such a high level that a few hours after peak daily flow (about 0.57 gpm/sq ft)

FIGURE 6-13: DAILY VARIATION IN ALUM PRIMARY SLUDGE LEVEL AND EFFLUENT SUSPENDED SOLIDS FOR CHEMICAL CONTACTOR

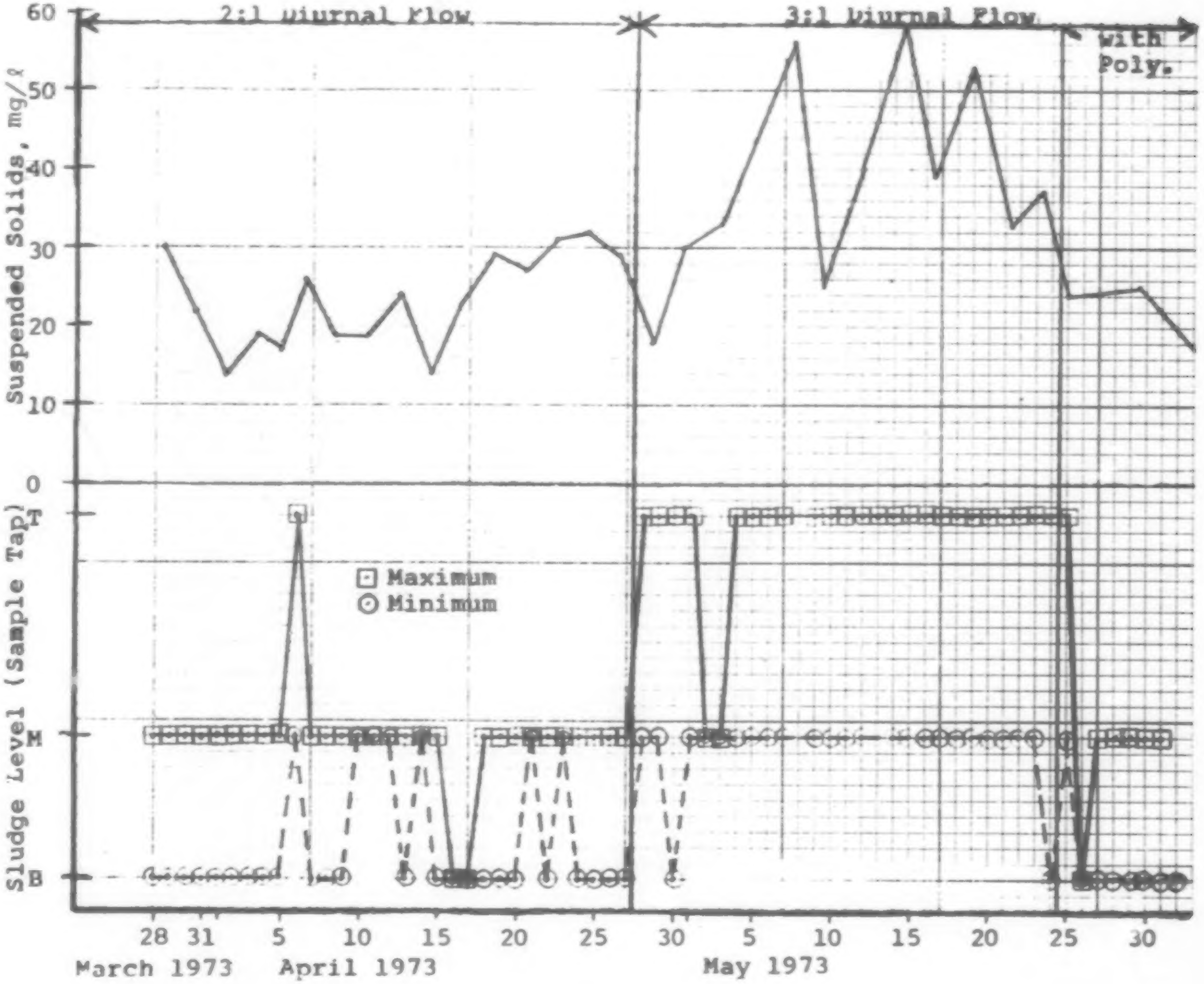


FIGURE 6-14: TYPICAL DIURNAL FLOW PATTERNS

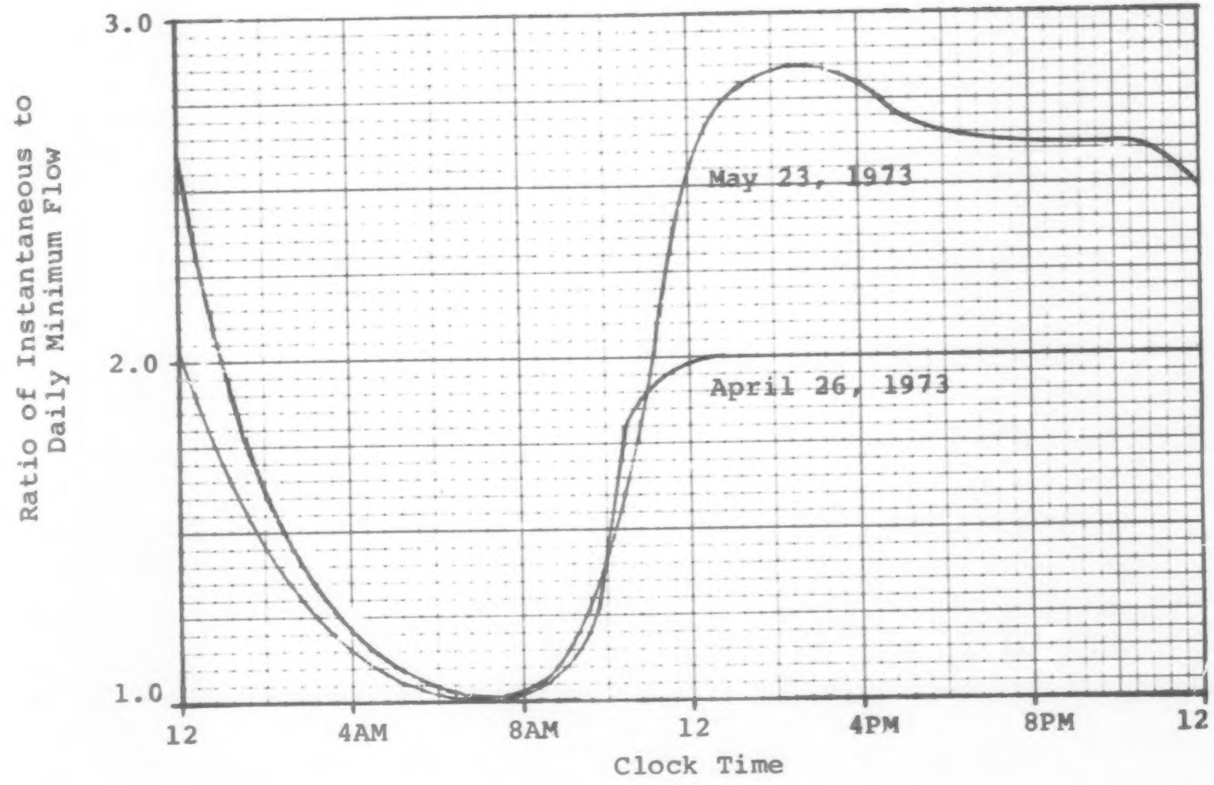
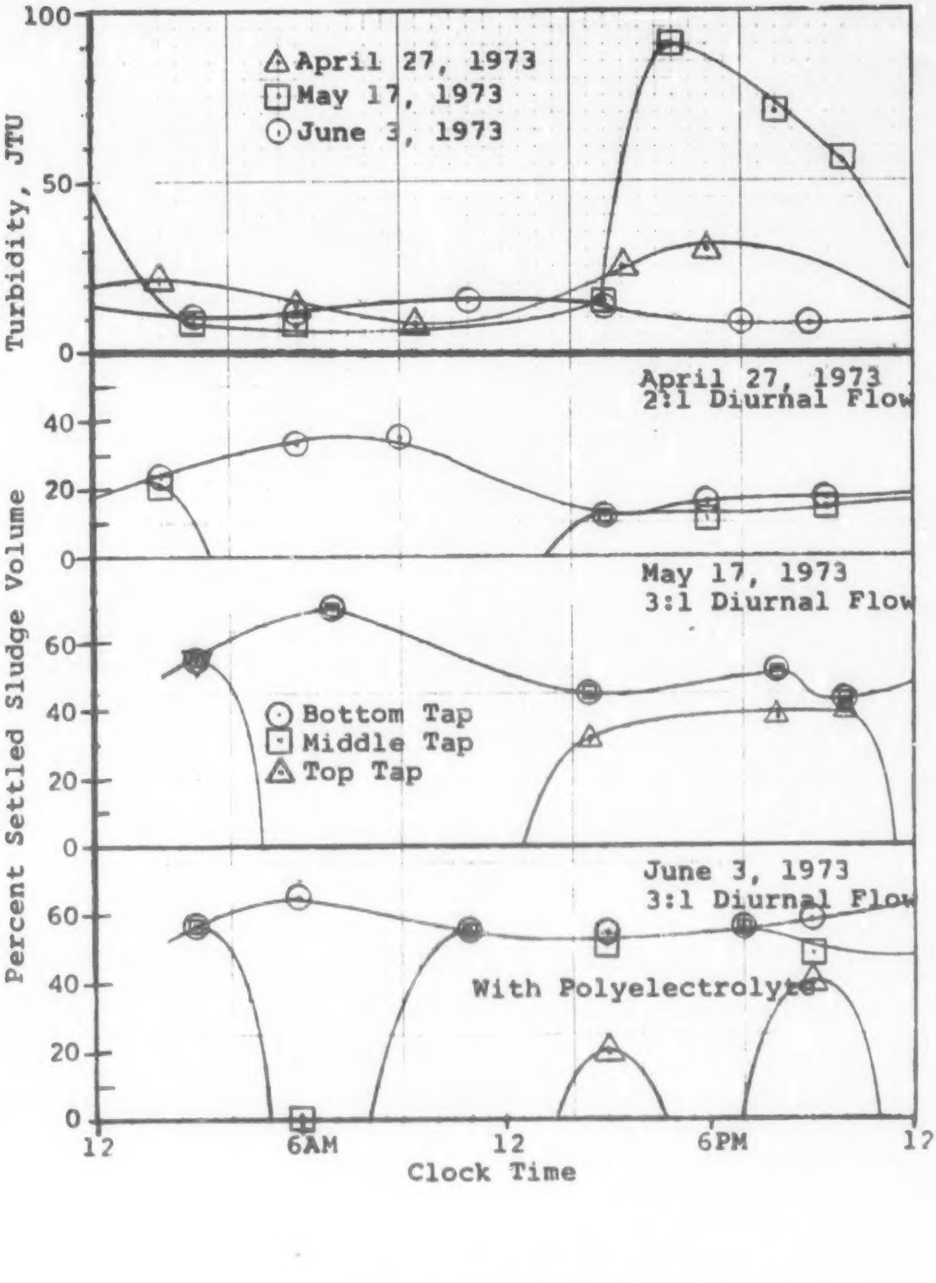


FIGURE 6-15: TYPICAL HOURLY VARIATIONS IN ALUM PRIMARY SLUDGE CONCENTRATION AND EFFLUENT TURBIDITY



the sludge-blanket rose and began to overflow the effluent launders. The basic cause of the problem was not a high flow, but rather poor operation. The sludge inventory (depth and concentration) had been allowed to become too great. On this specific day (May 17), the sludge pool interface rose in the clarification zone at a rate of about one ft/hr. Had operating personnel been aware of the high sludge inventory situation and had checked the units performance more than once during the immediately preceding 8 hour shift, the clarification failure might have been averted.

Inspection of Figure 6-15 shows that use of 0.25 mg/ℓ of anionic polyelectrolyte at the 3:1 diurnal flow resulted in an extremely stable sludge pool and effluent clarity. The variation in maximum and minimum daily sludge concentration (i.e., 10 minute percent settled sludge volume) is readily apparent in Figure 6-15. Figure 6-16 shows these variations over the entire operating period. The general effect of diurnal flow pattern and use of polyelectrolyte is apparent.

From the results presented in Figure 6-13 through 6-16 it can be concluded that neither effective or reliable clarification was achieved at a sustained peak daily hydraulic loading of 0.57 gpm/sq ft, without the use of polyelectrolyte. Effective and reliable clarification was achieved at a sustained peak hydraulic loading of 0.42 gpm/sq ft. These conditions are consistent with perviously reported results.[1]

On June 8, 1973, use of $FeCl_3$ in the chemical treatment solids-contact unit was commenced. A 3:1 diurnal flow with a peak daily hydraulic loading of 0.57 gpm/sq ft was attempted. After about a week of futile effort it was concluded that this was an excessively high peak hydraulic loading. The plant flow was changed to a 2:1 diurnal flow pattern with a peak hydraulic loading of about 0.45 gpm/sq ft and acceptable performance resulted.

The daily sludge pool interface depth variations were essentially the same as for alum treatment shown in Figure 6-13 (2:1 flow). Daily variations in sludge pool concentration (10 minute percent sludge volume) and effluent suspended solids values are shown in Figure 6-17. The effectiveness of polyelectrolyte in densifying and increasing the settling rate of the sludge and thus reducing concentration variations is obvious. As noted on Figure 6-17 and discussed for operating Period IV, the effluent suspended solids concentration shown included post-precipitation iron compounds. Thus it was not possible to determine the exact cause of the variable effluent suspended solids concentrations.

FIGURE 6-16: DAILY VARIATION IN ALUM PRIMARY SLUDGE CONCENTRATION AT BOTTOM SAMPLE TAP

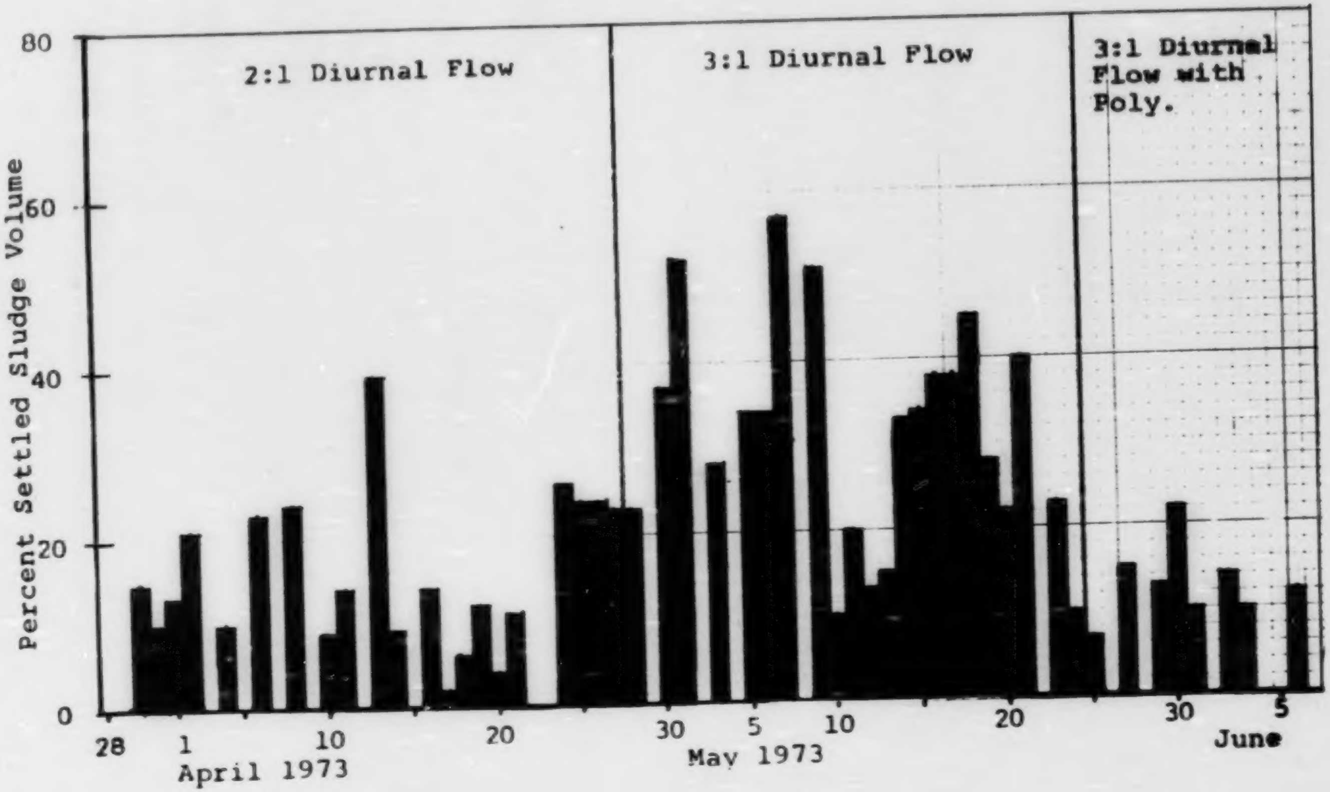
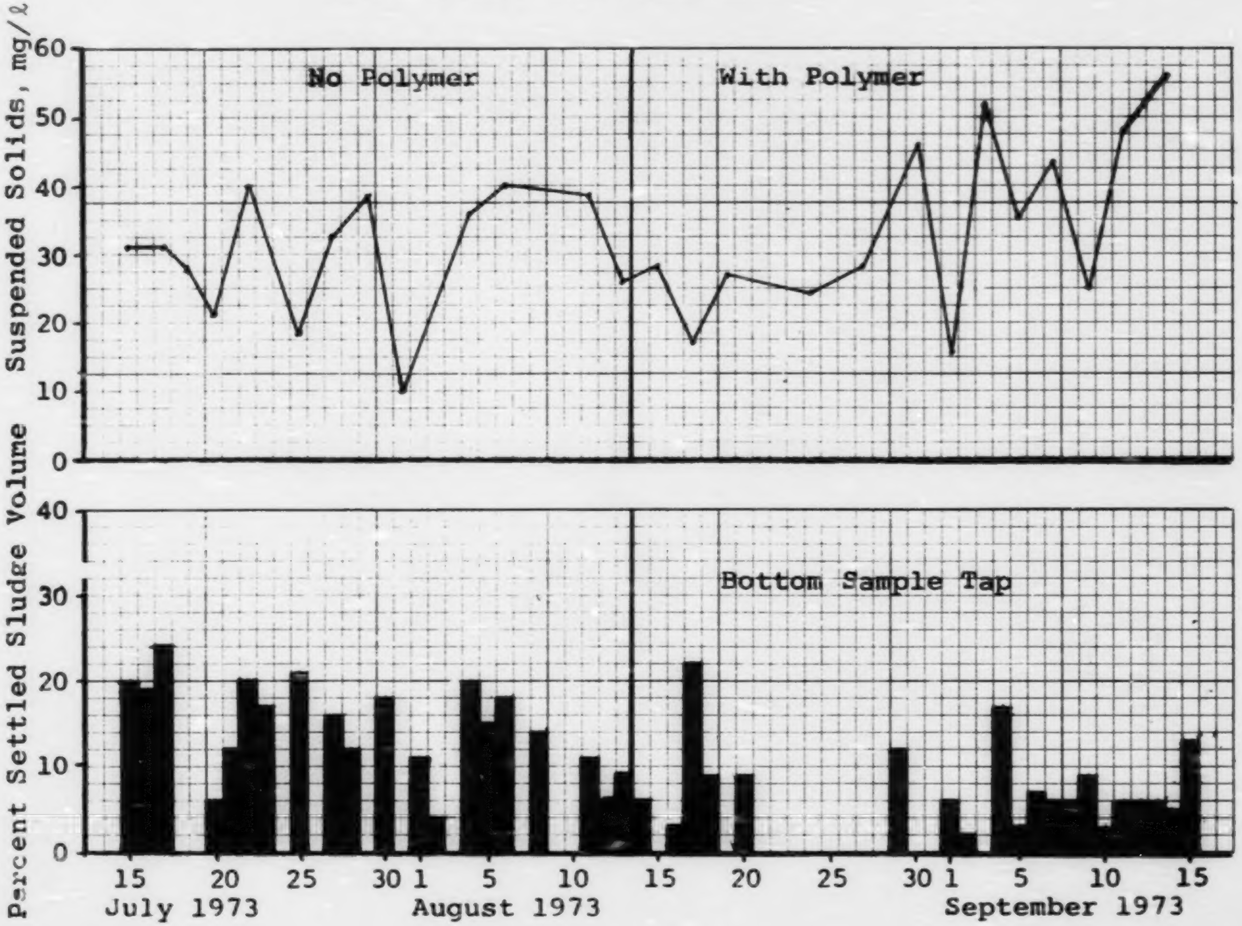


FIGURE 6-17: DAILY VARIATION IN FERRIC PRIMARY SLUDGE CONCENTRATION AND EFFLUENT SUSPENDED SOLIDS



Carbon Application

Solids-contact units were the heart of the carbon treatment system for soluble organic removal. The solids-contact units provided mixing, contact time, sludge inventory and rough liquid-solids separation. Unit processes performed were adsorption, anaerobic biological activity and gravity sedimentation. The clarification effectiveness of the solids-contact units under constant flow conditions had been extensively evaluated previously.[1]

To further evaluate the operational stability and clarification effectiveness of the carbon contactors, results for four various sets of operating conditions will be presented and discussed. Figure 6-18 shows daily sludge pool interface depth variations and effluent composite sample suspended solids concentrations for these four sets of conditions. Table 6-8 presents pertinent operating conditions. For peak: minimum diurnal flow variations up to 2.3:1, very stable sludge pools existed. Maximum to minimum sludge pool depth variations were 1 to 3 ft. At the higher diurnal flow variation of 2.9:1, the variation may have been slightly larger.

TABLE 6-8:

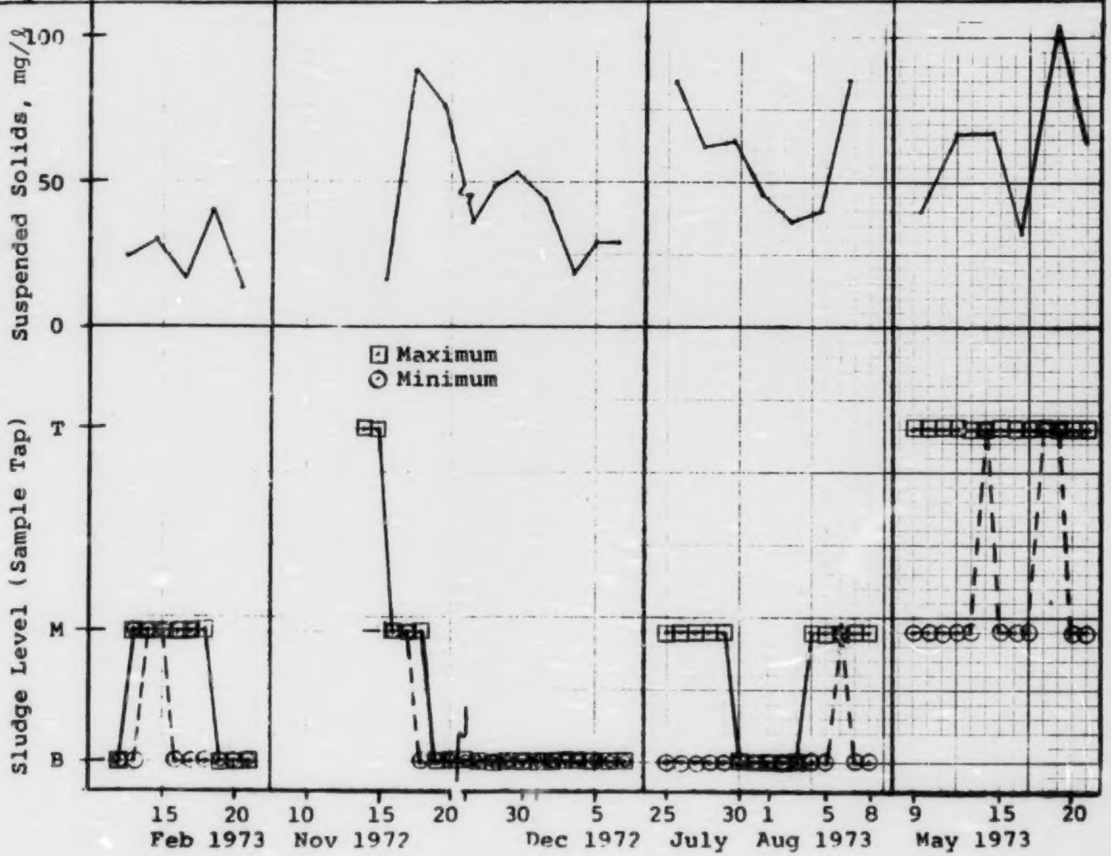
PERTINENT OPERATING CONDITIONS FOR PERIODS SHOWN IN FIGURE 6-18

Period Dates	February 1973	November December 1972	July August 1973	May 1973
Pretreatment Chemical	Lime	Lime	$FeCl_3$	Alum
Contacting[a)] Mode	SS	SS	2SCC	2SCC
Stage	--	--	1st	1st
Flows: gpm/sq ft				
Peak	--	0.90	0.62	0.79
Average	0.56	0.72	0.52	0.55
Minimum	--	0.45	0.27	0.27
Peak:Minimum Flow	1:1	2:1	2.3:1	2.9:1

a) SS-single stage; 2SCC-two stage counter-current

FIGURE 6-18: DAILY VARIATION IN CARBON SLUDGE LEVEL AND EFFLUENT SUSPENDED SOLIDS FOR CARBON CONTACTORS

Pretreatment Chemical	Lime	Lime	Ferric Chloride	Alum
Diurnal Flow	1:1	2:1	2.3:1	2.9:1
# of Stages	1	1	2 (1st stage)	2(1st stage)



The lowest variability and overall average effluent suspended solids concentration were experienced during the constant flow period. These data were not, however, directly comparable to the other periods shown since the granular media filter performance would contain carbon fines, and chemical floc, within the carbon-contacting clarification system, due to recycle of backwash to the carbon contactor. If these fines were to build up, or not report to the spent carbon sludge treatment system, clarifier effectiveness would be expected to diminish.

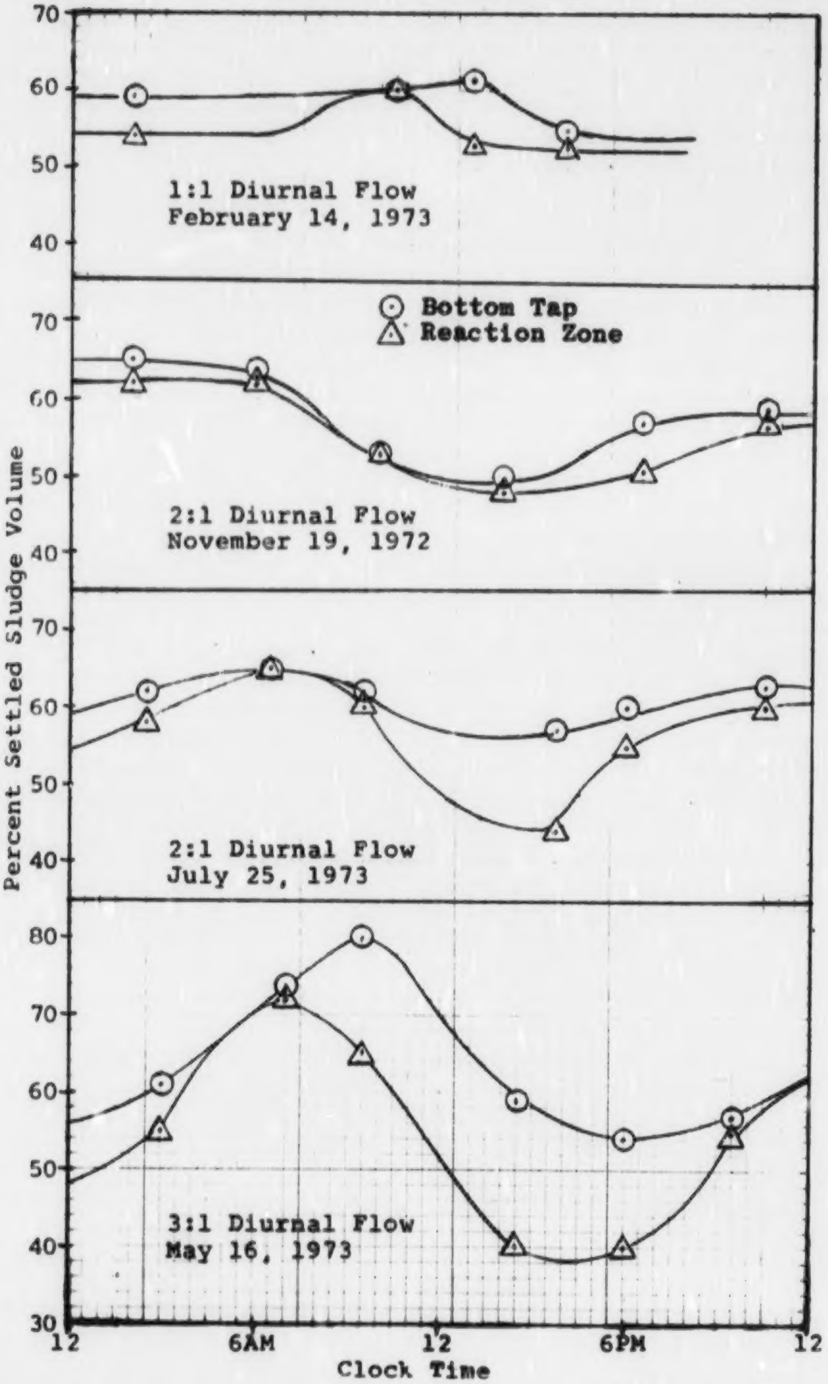
Except for the first few days during mid-November 1972, shown in Figure 6-18, when carbon inventory was being gradually reduced, reasonably good clarification was experienced. The variability of effluent suspended solids concentration during the higher two diurnal flow periods was not substantially greater than for the lower (2:1) diurnal flow period. Overall average effluent suspended solids concentrations were, however, significantly higher. Data in Table 6-8 indicate that peak hydraulic loadings were not excessive. During the July to August 1973, period when $FeCl_3$ pretreatment chemical was employed, some post-precipitation of iron may have occurred. At the peak hydraulic loading of 0.62 gpm/sq ft, efficient removal of iron floc would not have been likely. The high effluent suspended solids concentrations shown for the May 1973, period in Figure 6-18 may have been partly due to a pretreatment effect; namely, 30 to 50 mg/ℓ of suspended solids in the carbon-contactor feed. This chemical floc did not settle in the chemical clarifier at the peak hydraulic loading of 0.57 gpm/sq ft. Because of this, it is unlikely that the chemical floc would have settled in the carbon clarifier which was operating at a higher peak hydraulic loading of 0.79 gpm/sq ft.

In general, it would have to be concluded that the clarification efficiency of the carbon treatment solids-contact units was not impressive. It will be shown that a properly designed and operated filter station effectively removed relatively high carbon-contactor effluent suspended solids concentrations.

The sludge pool concentration uniformity and stability, for typical days during the four previously discussed operating periods, is shown in Figure 6-19.

The effect of increasing diurnal flow ranges on sludge pool concentration, which is inversely related to sludge volume, is apparent. It is also obvious that in general the reaction zone and sludge pool concentrations were approximately equal.

FIGURE 6-19: TYPICAL HOURLY VARIATION IN CARBON SLUDGE CONCENTRATION



That the single day data present in Figure 6-19 was indeed typical is verified by the data presented in Figure 6-20. Figure 6-20 shows daily differences in maximum and minimum 5 minutes percent settled sludge volumes ($\propto$ suspended solids concentration) for the four operating periods described in Table 6-8 and Figure 6-18. It is interesting to note that in December 1972, at a peak daily hydraulic loading of 0.90 gpm/sq ft and very low sludge inventory conditions, a very stable sludge pool existed.

Summary Discussion

The results of this study show that properly designed, sized and operated solids-contact treatment units are applicable to physical-chemical treatment of municipal wastewater. Potential problems due to the presence of biodegradable material and diurnal flow patterns can be averted. Prudent operation is, however, just as important as prudent design of the units.

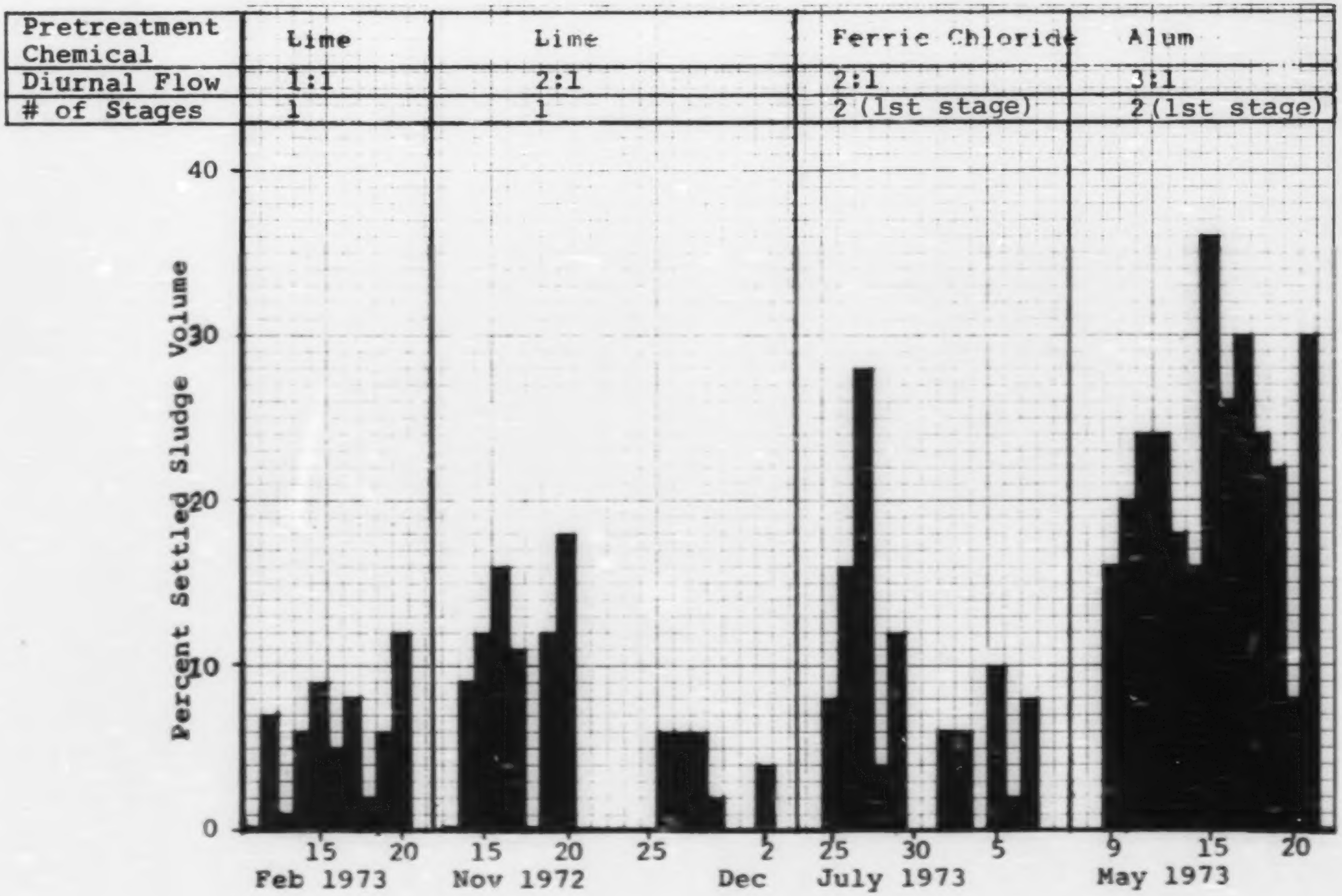
Of the applications evaluated in this study, carbon treatment was the most stable operation. It should be appreciated, however, that a high degree of clarification effectiveness by the carbon contactor-clarifiers was not a process requirement.

The results indicated that when polyelectrolyte was used with alum or $FeCl_3$ treatment the sludge depth and concentration variations were less than for lime treatment without polyelectrolyte. $FeCl_3$ treatment sludge depth and concentration variations were apparently less than for alum treatment sludge.

As with any gravity sedimentation unit the peak hydraulic loading is a critical design parameter. It should be appreciated that the critical rise rate in a solids-contact unit is as dependent upon the settling properties of the sludge pool as it is on the free settling properties of the feed suspended solids and precipitated material. Maximization of the chemical treatment clarifier underflow is very desirable. Maximization of hydraulic loading, at the expense of underflow sludge concentration may not minimize total treatment costs.

During this study a uniform volumetric rate of sludge blowdown was utilized. The results presented indicate that a non-uniform rate would have more likely produced a higher underflow solids concentration. Had sludge been blowndown only during the low flow period of diurnal flow, higher underflow suspended solids concentrations should have resulted.

FIGURE 6-20: DAILY VARIATION IN CARBON SLUDGE CONCENTRATION AT THE BOTTOM TAP



Pretreatment Chemical	Lime	Lime	Ferric Chloride	Alum
Diurnal Flow	1:1	2:1	2:1	3:1
# of Stages	1	1	2 (1st stage)	2 (1st stage)

Solids-contact units do not provide automatic treatment. They must be operated. An understanding of the chemistry, physics and hydraulics of the process, by the operator, is just as important as an understanding of the mechanical aspects, if not more so.

GRANULAR MEDIA FILTER PERFORMANCE

Previous contract experiences had indicated two problem areas concerning the filter performance and operation.[1] First, use of too fine a coal media resulted in only nominal penetration of suspended solids into the filter bed. The vast majority of suspended solids were removed in the upper few inches of coal resulting in excellent effluent quality but very short cycle times when high feed suspended solids were encountered. The second problem area previously experienced concerned backwashing. Because of the filter housing design (bed depth and free board) excessive backwash water was required (30 to 50 percent in excess of estimates). In addition, on one occasion a "mud ball" problem was experienced.

Based on the results of this previous study, it was recommended that evaluation of coarser coal media and the use of polyelectrolyte be made. It was also recommended that backwashing techniques be critically evaluated to define procedures which would prevent operational problems.

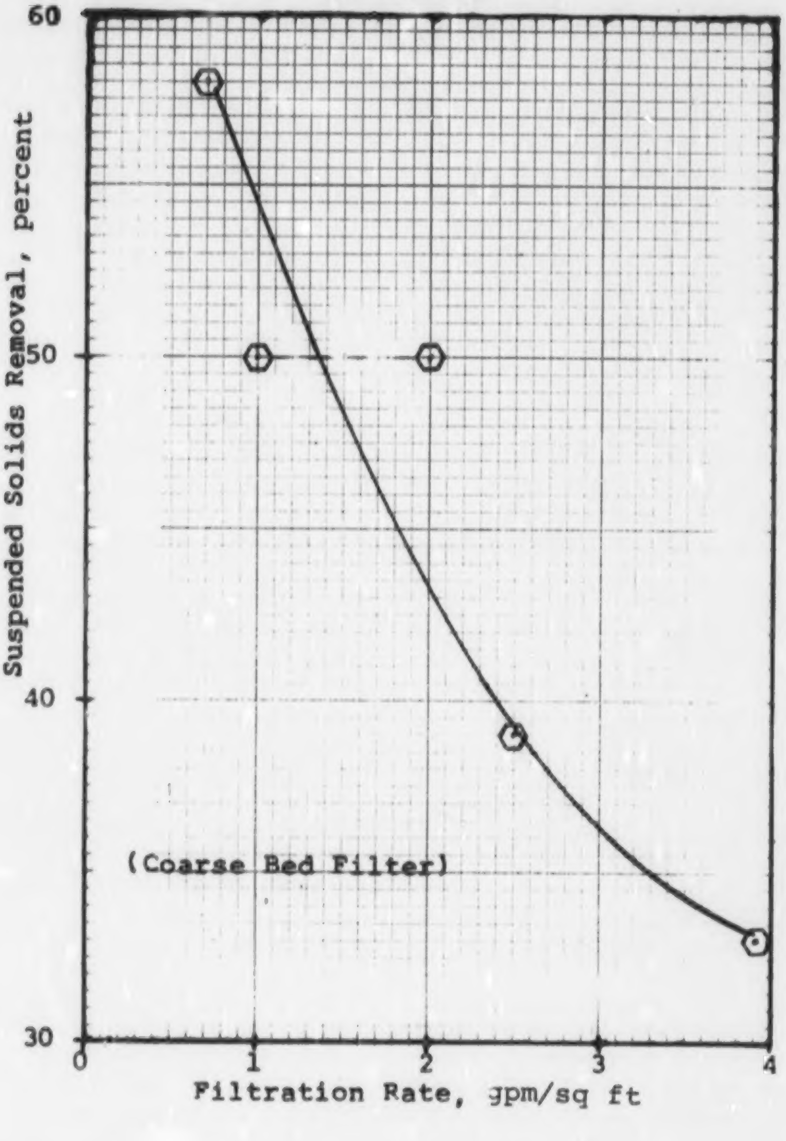
During the current study, two filter bed designs were evaluated. During Phase I a very coarse coal media was used and during Phase II a tri-media bed was used. Backwashing techniques were also characterized and the long term effectiveness evaluated.

Coarse Media Filter Bed

During the first 6 months of plant operation with lime pretreatment, a coarse bed filter was used. This filter bed consisted of 14 inches of 2.4 to 4.8-mm anthracite coal over 10 inches of 0.60 to 0.84-mm sand. It was anticipated that the sand layer would maintain a high quality effluent.

In general the suspended solids removal performance of the coarse bed filter was very poor. Figure 6-21 shows four monthly average percent suspended solids removals at four filtration rates. Monthly average effluent suspended solids concentrations ranged from 6 to 21 mg/ℓ. It is apparent from inspection of Figure 6-21 that as filtration rates increased performance deteriorated considerably. As will

FIGURE 6-21: EFFECT OF FILTRATION RATE ON SUSPENDED SOLIDS REMOVAL EFFICIENCE



be discussed later, use of anionic polyelectrolyte was found to improve performance substantially.

The suspended solids removal pattern in the filter bed was variable. Figure 6-22 shows typical terminal head loss patterns. At low filtration rates, some penetration of suspended solids into the coal-sand interface occurred. For the runs shown, about 50 percent suspended solids removal was achieved. Apparently most of the suspended solids that penetrated through the coal layer were not removed by the smaller sized sand media.

A series of filter runs were conducted, at constant filtration rates, to demonstrate the effect of using polyelectrolyte on the suspended solids removal pattern and performance. Results shown on Figure 6-23 indicate that at the higher polyelectrolyte dosage increased amounts of suspended solids were removed in the coarse coal layer. Without polyelectrolyte (Run #130), only about 20 percent removal of 24 mg/ℓ feed suspended solids was experienced. With polyelectrolyte, 50 to 70 percent removal of 17 and 20 mg/ℓ feed suspended solids was achieved. The effluents with 9 and 6 mg/ℓ of suspended solids were considered marginally acceptable.

The coarse bed filter was routinely backwashed using the following procedures:

1) Lower the water level by draining for 4 to 8 minutes to the surface of the coal.

2) Air-scour at about 3.5 SCFM/sq ft for 2 to 3 minutes.

3) Water wash at about 20 gpm/sq ft for 5.5 to 7.5 minutes.

A higher water backwash rate would have been desirable, but limited pumping capacity made it impossible. Bed expansion during water backwash was only about 10 percent (@ 13 to 16°C). Actually the coal layer was not observed to expand significantly, whereas the sand layer expanded by about 20 percent. The coal media was not fluidized, but relative rolling of filter media grains was observed. Apparently the coal media was coarse enough to allow flushing of suspended solids removed in the sand layer.

Use of the above backwash procedure for over six months did not result in any "mud ball" formations or long-term buildup in clean bed head loss (i.e., immediately after backwashing). It was concluded that this backwash procedure was effective

FIGURE 6-22: TERMINAL HEAD LOSS PATTERN FOR COARSE BED FILTER

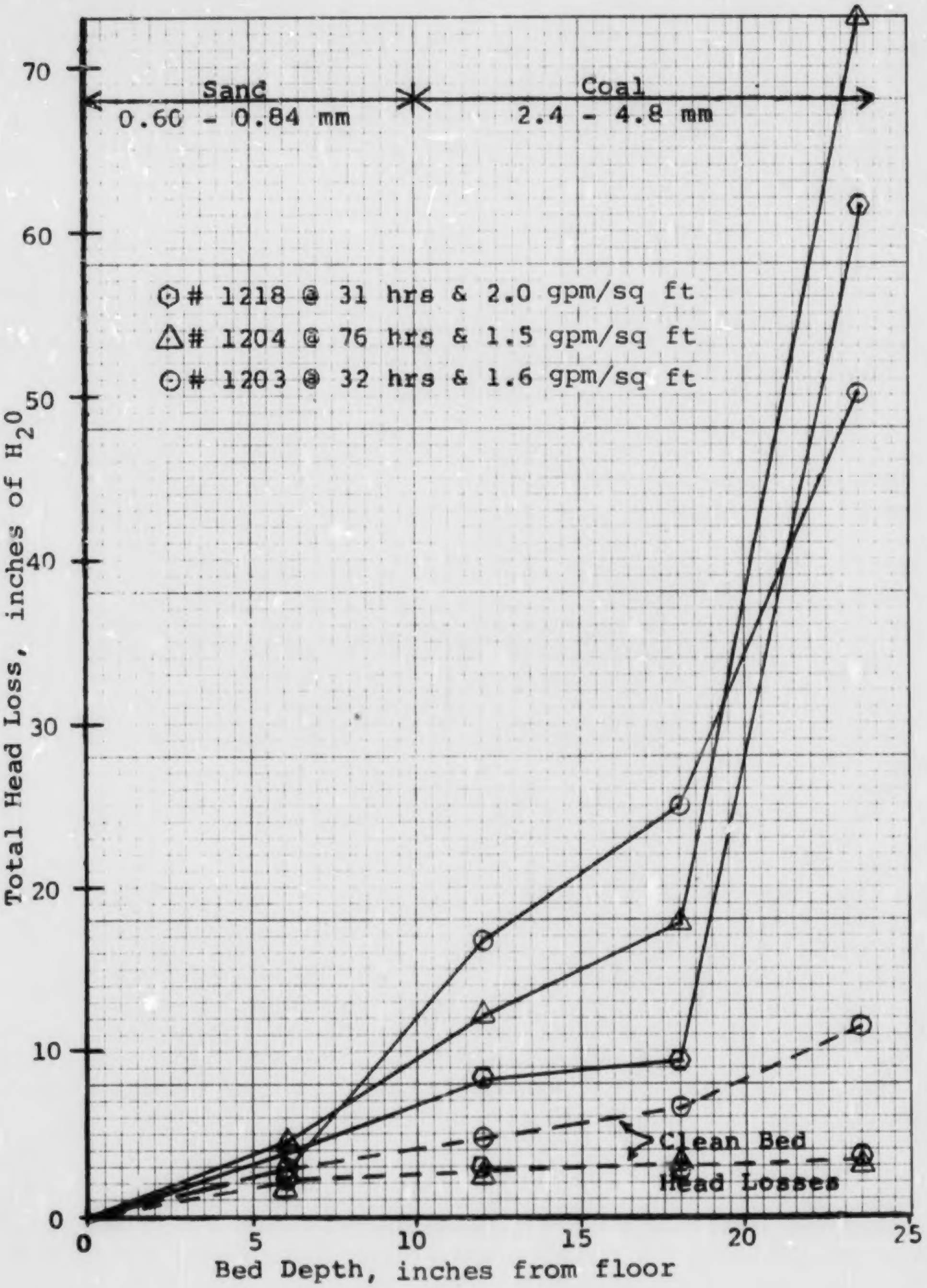
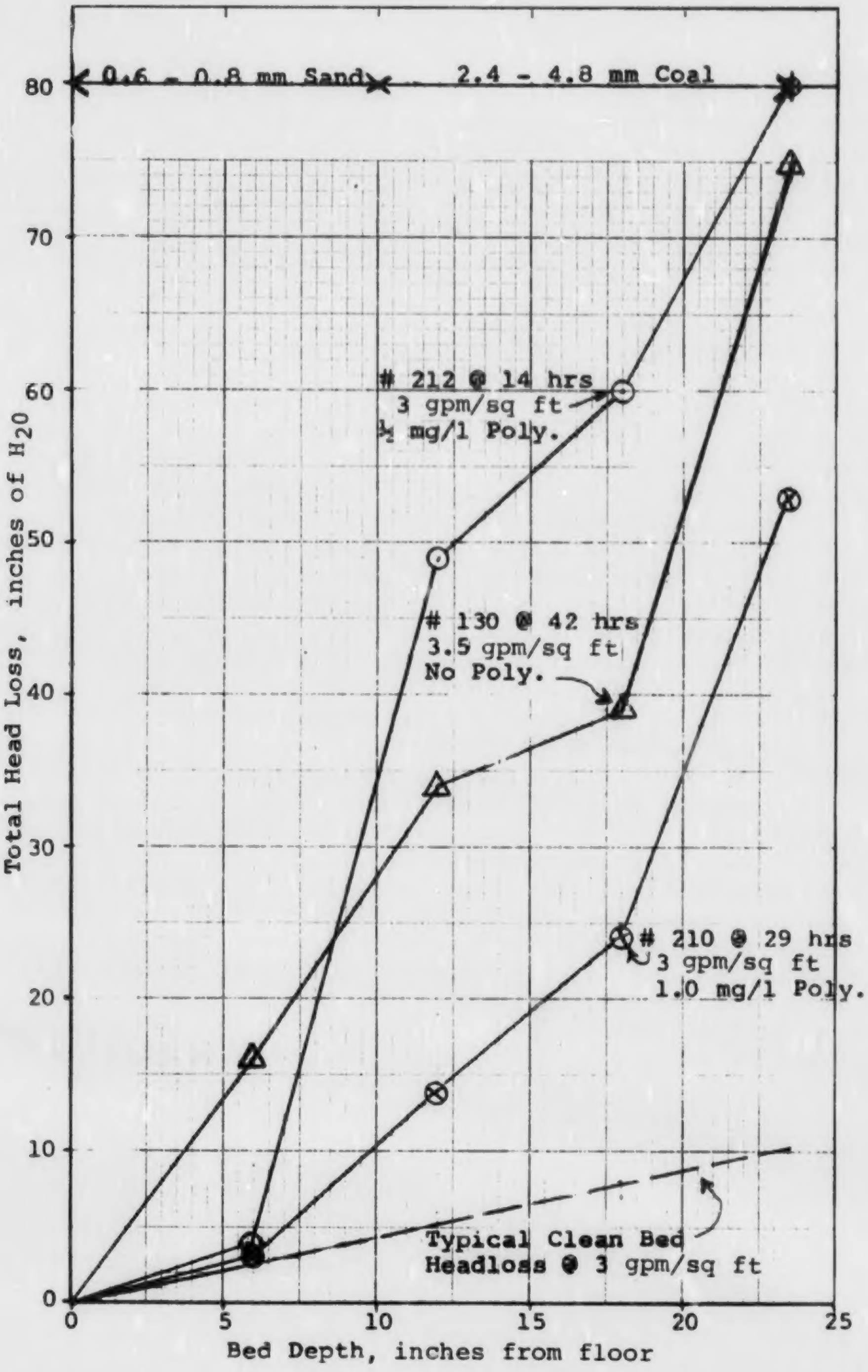


FIGURE 6-23: EFFECT OF POLYELECTROLYTE ON TERMINAL HEAD LOSS PATTERN



for the filter bed used. For a 2 gpm/sq ft filtration rate the above water backwash condition resulted in recycling less than 5 percent of the filtered water for cycle times greater than or equal to 21 hours.

Tri-Media Filter Bed

Having evaluated two extremes in the size of coal for the dual-media filter beds used during previous and current contract studies, it was decided to try a medium sized coal. Results for the coarse media filter bed design indicated that the 0.60-mm top size sand had not consistently produced the desired effluent suspended solids concentration of 5 mg/ℓ.

To prevent gross mixing of coal and sand layers there was a limit to the minimum sand top size which could be used. Prevention of gross intermixing at normal water backwash rates, requires that the largest coal size be no more than about 2.5 times the smallest sand size. As a result of these limitations use of a third layer of garnet sand (density 4.2 g/cc) was considered necessary to have a layer of media in the filter bed with a particle size considerably less than 0.6-mm. A tri-media (or multi-media) filter bed was designed and installed resulting in the approximate media depths and sizes shown in Figure 6-24. Each layer of media was made of commercially available cuts which were backwashed at about 30 percent bed expansion and then scalped. About 11 inches of standard 60 to 80 U.S. Mesh Idaho Garnet* was installed, backwashed three times and 3 inches (about 27 percent) scalped off. One thousand lbs of 20 by 30 U.S. Mesh filter sand was then installed. After backwashing at about 30 percent bed expansion considerable intermixing of the garnet and sand was observed. There was, however, a distinguishable layer of garnet of about 6 inches. About 4.5 inches of sand was skimmed off. About 25 inches of anthracite coal+ (1.2 to 1.5-mm effective size, <1.7 uniformity coefficient) was then installed, backwashed three times to a total bed expansion of about 30 percent and 5 to 6 inches scalped off. Slight intermixing at the coal-sand interface was observed. The resulting filter bed, as shown in Figure 6-24, was used for the last 7 months of this study except that in late April about 5 to 6 inches of coal was lost during an unattended backwashing.

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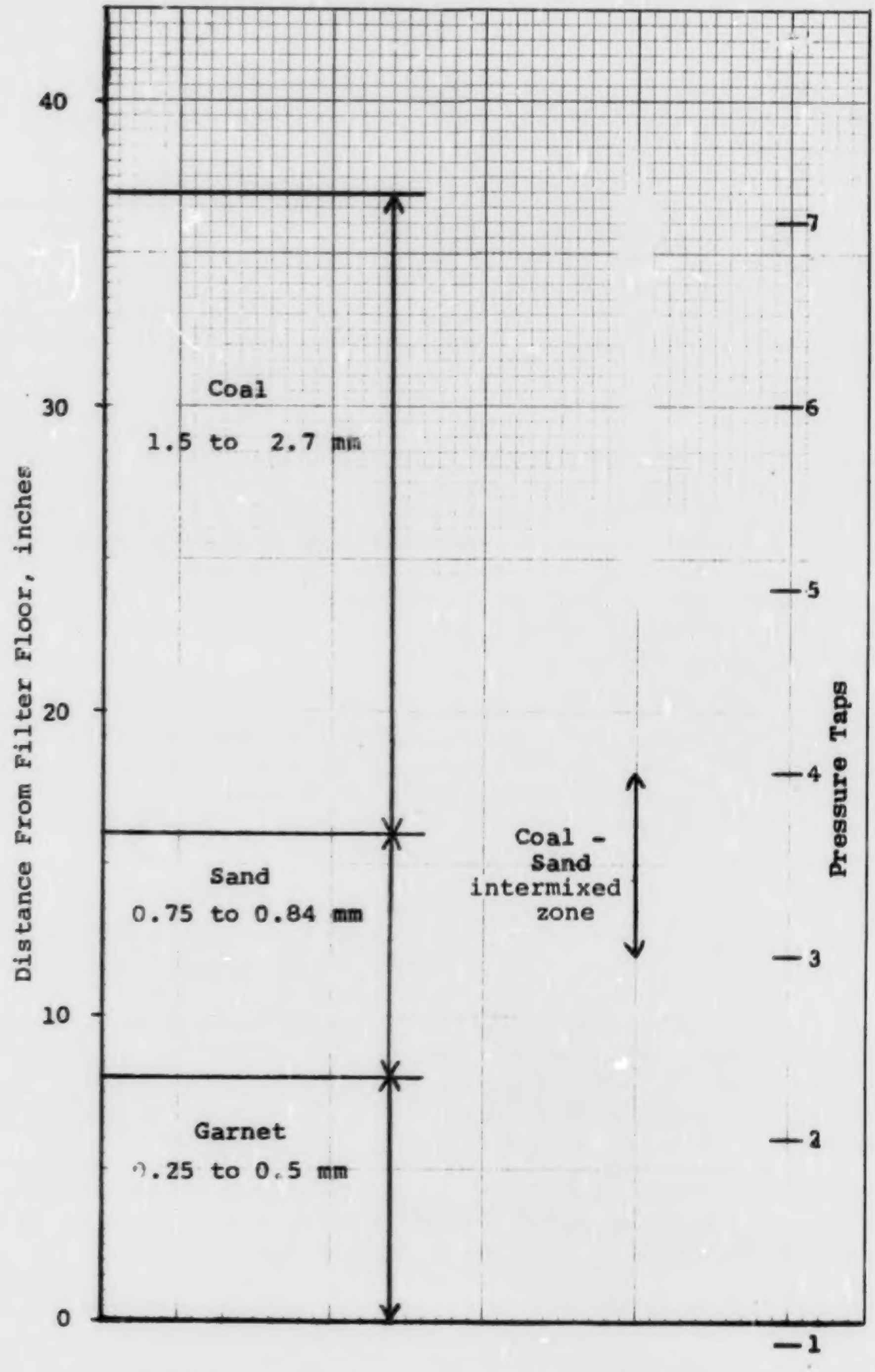
* A product of Idaho Garnet Abrasive Company, Kellogg, Idaho

+ A product of Reading Anthracite Coal Company, Pottsville, Pennsylvania

FIGURE 6-24: TRI-MEDIA FILTER BED DESIGN


40
30
20
10
0
Distance From Filter Floor, inches
Coal
1.5 to 2.7 mm
Sand
0.75 to 0.84 mm
Garnet
0.25 to 0.5 mm
Coal - Sand intermixed zone
Pressure Taps
7
6
5
4
3
2
1

It should also be noted that on February 13, 1973, the original all-plexiglass filter housing was fractured beyond repair. A new steel filter with a 2 ft wide plexiglass window over the full bed depth was fabricated, installed and started-up on April 13, 1973.

The following sub-sections discuss the performance of the tri-media filter during the last four operating periods (April to November 1973).

April 1973: During April 13 to 28, 1973, the pilot plant feed flow was as depicted in Figure 6-14, at a 2:1 peak to minimum diurnal flow. Due to substantial effluent sample flows from the three preceding solids-contact-clarifiers (about 13 gpm constant flow) the actual flow variation to the filter was much higher. The approximate peak, average and minimum daily filtration rates were 3.0, 2.4, and 0.6 gpm/sq ft, respectively.

At these filtration rates and for the other plant operating and performance conditions (see Table 6-3), the filter suspended solids removal performance was good. An average suspended solids removal of 90 percent was achieved resulting in an average effluent suspended solids of 2.8 mg/ℓ (1 to 4 mg/ℓ range). At a terminal headloss of 6 to 7 ft of water, filter cycle times of from 4 to 20 hours with an 8 hour average was experienced.

Figure 6-25 shows the typical, normalized total filter headloss development pattern experienced for this period of operation. The very rapid increase in headloss buildup near the end of the cycle is indicative of a suspended solids removal pattern where a majority, or all, or the filter media pores at some given depth are completely clogged.

Figure 6-26 shows three typical terminal headloss versus bed depth patterns observed. Two features of the patterns shown were judged desirable. First, few suspended solids were removed in the bottom 6 to 12 inches of the filter bed, thus implying a low suspended solids concentration in the filter effluent. Secondly, substantial removal of suspended solids occurred in both the top half of the coal layer and the top few inches of the sand layer as indicated by the substantial headloss at these locations. Since the void sizes in the coal layer are substantially larger than in the sand layer, the near equivalent headloss observed implies that substantially more suspended solids were removed in the coal layer than in the sand layer.

FIGURE 6-25: TYPICAL NORMALIZED HEAD LOSS BUILD-UP PATTERN

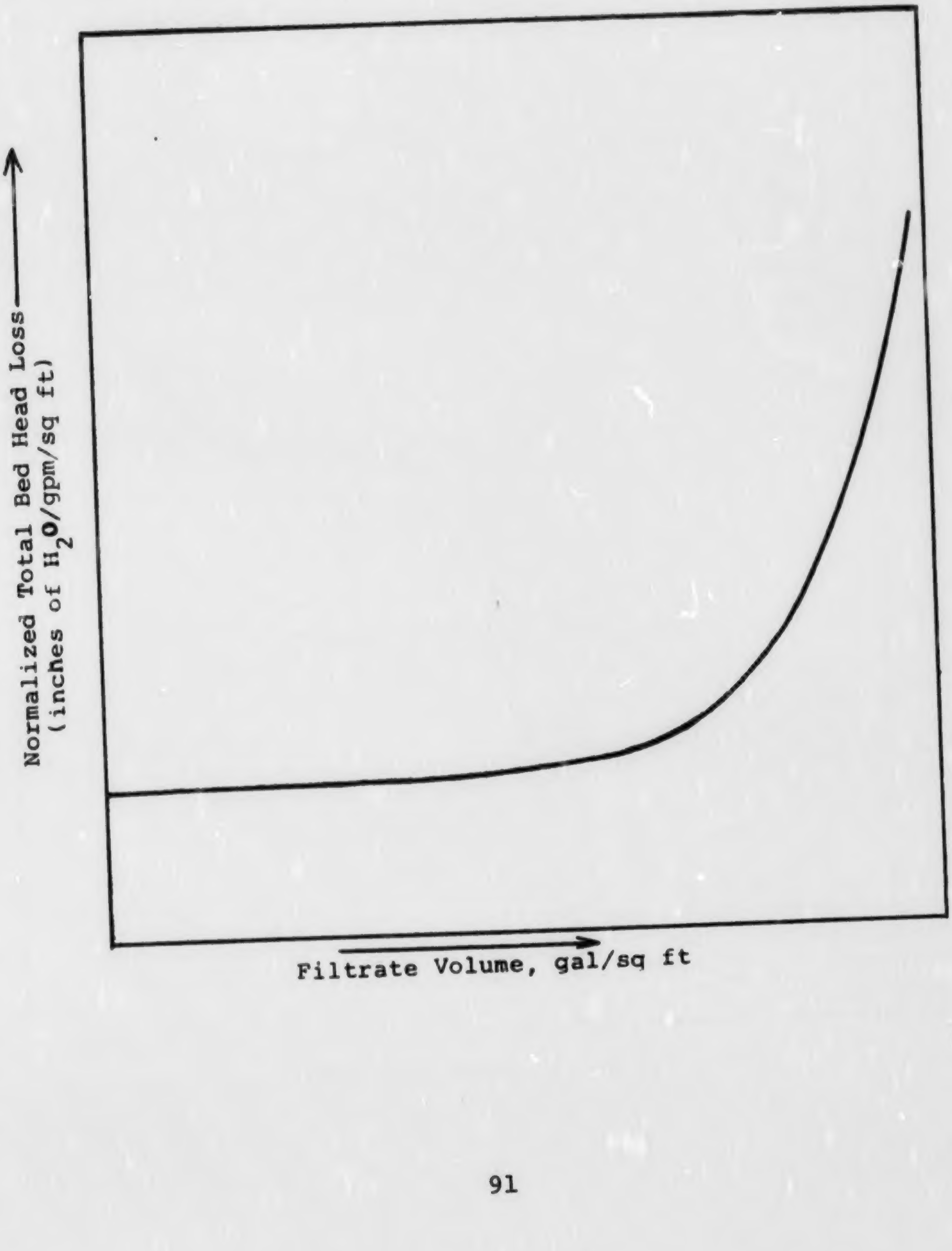
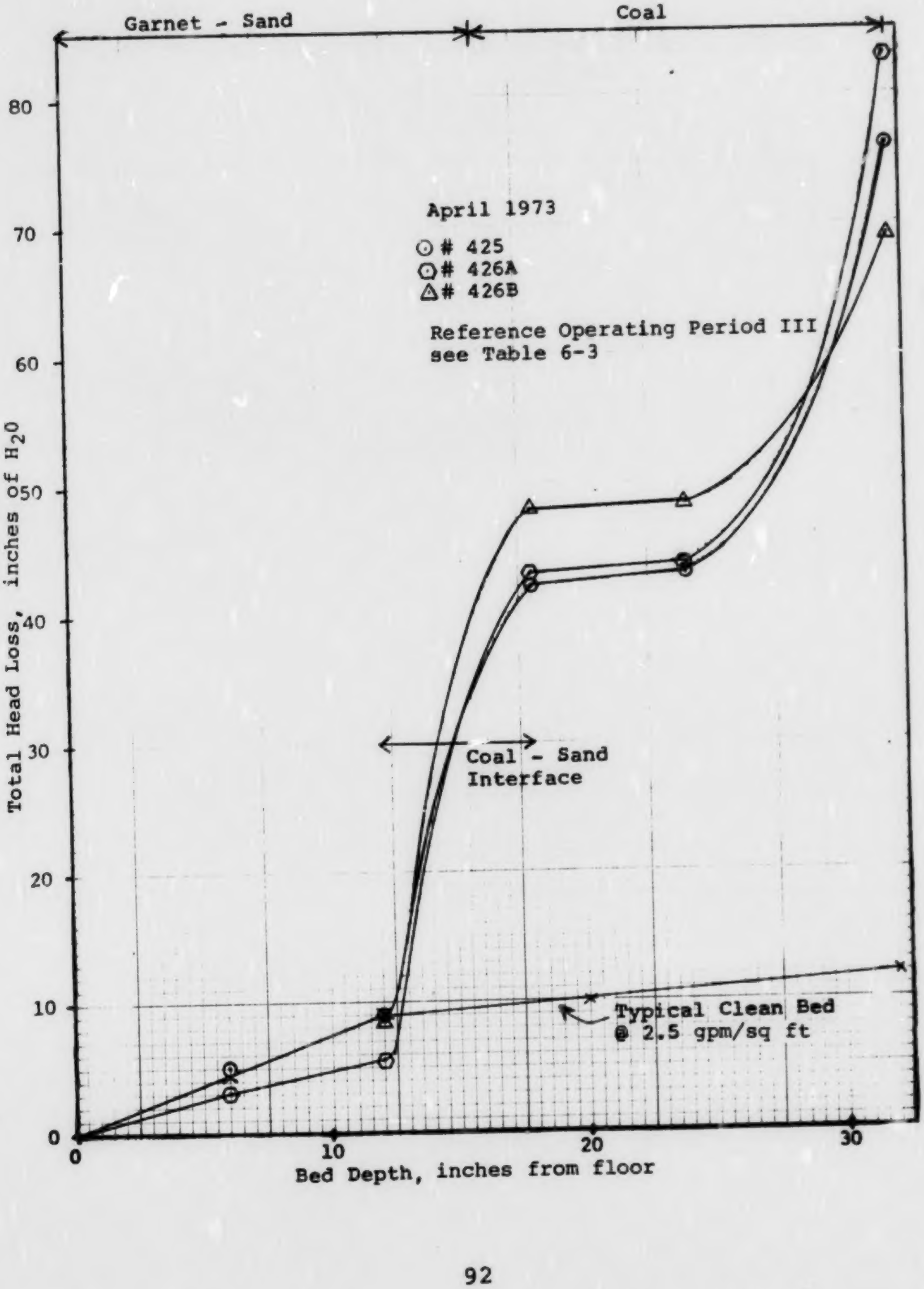


FIGURE 6-26: TERMINAL HEAD LOSS VERSUS BED DEPTH PATTERN



The terminal headloss patterns shown in Figure 6-26 supports the inferred pore clogging suspended solids removal pattern. As noted above, the substantial suspended solids removal at the top few inches of sand suggests that all voids in the top coal layers were not clogged (i.e., substantial suspended solids penetrated into the sand layer). During this operating period a few filter runs were terminated, at less than 6 to 7 ft headloss, due to breakthrough of suspended solids into the effluent. These breakthroughs typically occurred during later morning when the filtration rate would increase by a factor of 5, from a minimum of about 0.6 to a maximum of 3 gpm/sq ft, over a period of 4 to 5 hours. These breakthroughs were graphically indicated by the continuous recording of effluent turbidity as measured by a Hach CR-low range nephlometer unit.*

The tri-media filter bed was routinely backwashed using the following procedure:

1) Lower the water level by draining for 8 to 15 minutes to the surface of the coal.

2) Air-scour at about 3.5 SCFM/sq ft for 1.5 to 2.5 minutes.

3) Water backwash at about 30 gpm/sq ft for 2.8 to 1 minute.

The average backwash used about 110 gal/sq ft. At an average filtration rate of 2.4 gpm/sq ft and a cycle time of 8 hours, this resulted in a somewhat high recycle of 6.8 percent of the filtrate. At 30 gpm/sq ft and about 17°C water temperature, the total filter bed expansion was about 30 percent.

On April 24, 1973, about 5 inches of coal media was flushed out of the filter with the backwash water. The cause apparently was due to a substantial depth of the coal layer rising as a plug to the backwash outlet. In the senior author's experience this was an anomalous observation for a 3.5 ft diameter pilot plant filter.

The recycle of lost coal media to the carbon-contacting system and subsequently to the spent carbon thickeners resulted in the nuisance of frequent underflow pump clogging for several days.

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*Manufactured by Hach Chemical Company, Ames, Iowa

May 1973: During the last half of plant operating Period III, the peak and average plant flows were increased to increase the diurnal variation to about 3:1 peak:minimum flow. This resulted in the filter experiencing about a 6:1 peak:minimum diurnal flow variation. Of significance also was the doubling of filter feed suspended solids due to the higher overflow rate in the carbon-contactor-clarifiers.

During the period April 28 to May 6, 1973, the filter was subjected to peak, average and minimum daily filtration rates of about 4.6, 2.9 and 0.7 gpm/sq ft. Under these operating conditions, only 70 to 80 percent removal of an average feed suspended solids of about 60 mg/ℓ was achieved. Filter cycle times were erratic, varying from 1.5 to 15 hours. Breakthrough of suspended solids into the effluent was the rule rather than the exception. Terminal filter headloss patterns indicated little removal of suspended solids in the coal layer. The vast majority of terminal headloss existed at the coal-sand interface.

On May 6, 1973, the filter flow was cut in half resulting in peak, average and minimum daily filtration rates of 2.3, 1.5 and 0.4 gpm/sq ft. Feed suspended solids remained about 60 mg/ℓ (40 to 100 range). Suspended solids removal performance remained somewhat erratic in the 75 to 85 percent removal range. The majority of headloss buildup continued to occur at the coal-sand interface.

During the period May 10 to 15, 1973, several filter runs were conducted adding about 1/4 mg/ℓ anionic polyelectrolyte just ahead of the filter. Effluent suspended solids concentrations were reduced by about half. Terminal filter bed headloss patterns were somewhat erratic but generally indicated substantial suspended solids removal in the coal layer. No real breakthrough of suspended solids was experienced. Very short filter cycles of just a few hours (2 to 5) were typical.

On May 21, 1973, polyelectrolyte feed was started to the chemical treatment unit in an effort to obtain lower effluent suspended solids and more stable operation. A slight improvement in filter suspended solids removal performance was experienced. Filter cycle times increased to a range of 17 to 25 hours. Terminal filter bed headloss data indicated that most suspended solids were still being removed at the coal-sand interface.

In general it was concluded that operation of the pilot plant under the stress of relatively high flows used during May caused relatively poor filter performance. Use of polyelectrolyte just ahead of the filter did result in a more uniform effluent quality (i.e., no breakthrough of suspended solids), but prohibitively short cycle times resulted.

June-September 1973: During operating Period IV and V (6/10/73 to 9/14/73) the performance and operation of the filter was exceptionally good. Part of the reason was the nominal peak, average, and minimum daily filtration rates of about 3.3, 2.5, and 1.0 gpm/sq ft.

During Period IV (see Table 6-4) no polyelectrolyte was added. Chemical treatment was achieved using $FeCl_3$. Feed suspended solids to the filter averaged 65 mg/ℓ (20-120 range) and the effluent suspended solids averaged 3 mg/ℓ (1-11 mg/ℓ). This amounted to an average of 95 percent removal.

Filter cycle times ranged from 9 to 38 hour and averaged 23 hours. Terminal total filter bed headloss was in the 5 to 6 ft of water range.

The majority of suspended solids removal occurred at the coal-sand interface as indicated by the major headloss buildup at this location. Figure 6-27 shows the typical terminal headloss pattern experienced during Period IV. Twenty-four out of the 30 terminal headloss patterns observed exhibited this major headloss buildup at the coal-sand interface. Though penetration of suspended solids to the interface was the rule, few filter runs were terminated due to solids breakthrough into the effluent.

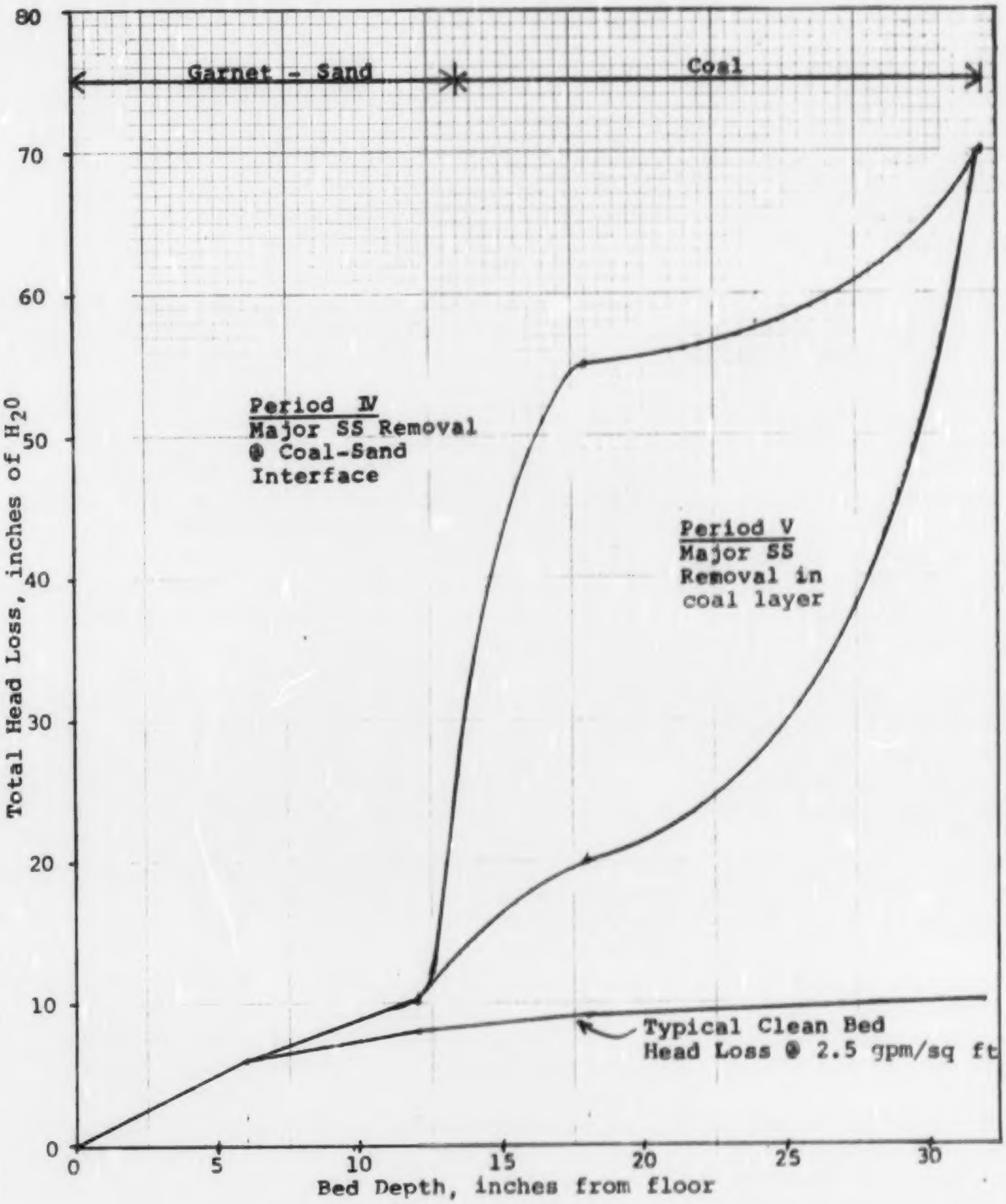
During Period V (see Table 6-5) polyelectrolyte was used with $FeCl_3$ chemical treatment. Filter feed suspended solids averaged 76 mg/ℓ (50 to 100 range) and effluent suspended solids averaged 4 mg/ℓ (1 to 8 mg/ℓ). This resulted in an average of 95 percent removal.

Filter cycle times ranged from 10-38 hours and averaged about 23 hours. Terminal total filter bed headloss was in the range of 5 to 6 ft of water.

During Period V the majority of suspended solids were removed in the coal layer as indicated by the major headloss buildup at this location. Figure 6-27 shows the typical terminal headloss pattern indicative of this situation.

Of some 20 terminal headloss patterns observed, 17 showed this pattern. It can be concluded that the use of polyelectrolyte in the chemical treatment unit resulted in a more readily "filterable" floc appearing in the filter feed. This observation emphasizes the fact that consideration of

FIGURE 6-27: TYPICAL HEAD LOSS VERSUS BED DEPTH PATTERN



the "total" treatment system must be taken into account when evaluating the effectiveness of any given unit operation.

Backwashing procedures used during Periods IV and V were identical to those noted previously for Period III. The use of about 100 gal of backwash water per sq ft when experiencing average filtration rates and cycle times of 2.5 gpm/sq ft and 23 hours, resulted in recycle of about 3.2 of the filtrate.

During one backwashing, when the total bed expansion was observed to be 22 percent, the coal layer was observed to have expanded about 17 percent and the garnet-sand layer 25 percent. The water fluidization properties of the two layers of media were obviously similar.

The backwash procedure used was judged to be effective in as much as no "mud ball" problems or increase in clean bed headloss pattern were experienced during this 3 month period of operation.

During previous contract work (Reference 1) using $FeCl_3$ treatment, a substantial coating of the coal media with a reddish-brown (iron compound) precipitate was observed. No such coating was observed during the current study.

September to November 1973: This last plant operating Period, IV, was from September 28 to November 8 during which time alum and polyelectrolyte were used in the chemical treatment step. Regenerated and reused carbon was used in the carbon contacting step. The peak, average and minimum daily filtration rates were approximately 4.7, 3.5, and 1.1 gpm/sq ft.

At the higher filtration rates, suspended solids removal performance deteriorated somewhat. Filter feed suspended solids averaged 38 mg/ℓ (16 to 109 mg/ℓ) and the effluent suspended solids averaged just under 8 mg/ℓ (3 to 19 mg/ℓ). The average percentage removal was thus only 79 percent. Performance was judged to be fair to good at the higher filtration rates.

Filter cycle times ranged from 8 to 31 hours and averaged about 18 hours. Terminal total bed headloss was typically in the range of 5 to 6 ft of water. Under these operating conditions approximately 2.9 percent of filtered water was used for backwashing and then recycled back to the first stage carbon contactor.

Typically most suspended solids were removed in the coal layer. Of some 19 terminal headloss patterns observed, 13 showed patterns similar to that indicated for Period V in Figure 6-27. Five out of the 19 observations indicated more uniform distribution of terminal headloss as shown previously in Figure 6-26. Comparison of suspended solids removals patterns for Periods III and VI, and considering differences in filtration rates, indicates that use of polyelectrolyte in the chemical treatment unit resulted in a more readily "filterable" floc appearing in the filter feed.

Backwashing of the filter media continued to be most effective. After 7 months of operation, no "mud balls", media fouling or change in clean bed headloss were observed. At the end of this study, the media was removed from the filter. It appeared to be quite clean and free from foreign material. No stagnant zones or "dead spots" were found in the filter bed.

Summary Discussion

The coarse coal filter bed evaluated during Phase I did not produce acceptable performance. The generally poor suspended solids removals and tendencies for solids breakthrough were unacceptable. The use of polyelectrolyte just ahead of the filter improved performance to a marginally acceptable level. The amount of polyelectrolyte needed, however, would result in a prohibitive cost.

In general the performance of the tri-media filter bed was rated good to excellent. At relatively high filtration rates and under certain pretreatment conditions performance deteriorated. Use of polyelectrolyte in conjunction with inorganic coagulants (alum or $FeCl_3$) in the chemical treatment step resulted in a more readily "filterable" suspension. Slightly higher filtration rates should be possible when using polyelectrolyte in the chemical treatment step.

Performance and operational results indicated that the "floc strength" of filtered suspended solids was fairly weak. Breakthrough of suspended solids at less than terminal headloss and at increasing filtration rates were partial indications of a "weak floc". Because of these observations it would be strongly recommended that any filter station, for this application, be designed to minimize sudden or drastic changes in filtration rate.

Based on the results of this study, the authors recommend that a tri-media filter bed, or a dual media filter bed with slightly smaller sand media be used. Use of polyelectrolyte in the chemical treatment step is also recommended, especially if inorganic coagulants are employed. Peak filtration rates should be restricted to less than 4 gpm/sq ft, and, if possible, an average filtration rate of 2.5 gpm/sq ft is suggested. Facilities for feeding about 1/2 mg/ℓ anionic polyelectrolyte in-line just ahead of the filter should be provided for but only used if needed to protect plant effluent quality. Auxiliary air-scour prior to water backwashing is also strongly recommended.

The granular media filter is the last liquid-solids separation step in the PAC-PCT flowsheet. It must continuously and reliably produce a highly clarified effluent. Conservative sizing is fully justified in the opinion of the authors.

SLUDGE TREATMENT

This section presents results and discussions of pilot plant and laboratory sludge dewatering studies. Gravity thickening and vacuum filtration of the three chemical-primary sludges and spent carbon sludge are evaluated.

GRAVITY THICKENING

Routine daily data collection around the gravity thickening units entailed determination of feed and overflow suspended solids concentrations, overflow and underflow volumes and average thickener sludge depth. Underflow suspended solids concentrations were determined on an intermittent basis, whenever the thickened sludge inventory and storage tanks were full. The total pounds of thickened sludge removed over a period of time (days) was determined. Thickener solids capture was consistently in the high 90 percent range, typically 99 percent. Thickener solids loading (TSL) in lb of dry solids per day per sq ft of thickener area was, therefore, determined on the basis of underflow solids only. Thus TSL values reported are an average for a fraction of a day to several days. The observed TSL, and thickened feed and underflow suspended solids varied significantly from average values during periods of inventory adjustments to the chemical treatment unit and/or thickener. Typical sludge inventory adjustment periods lasted no more than one to two days.

The approach used to identify pilot plant thickener performance was to plot feed and underflow suspended solids concentration, sludge depth and average TSL for each day of operation for all sludges. Periods of reasonably stable thickener operation (TSL and sludge depth) were identified. For these stable operation periods, average pertinent chemical treatment unit and thickener operating conditions and thickener performance data are reported.

The method of conducting and the approach to analyzing the results of laboratory thickening tests are contained in Appendixes A and B, respectively. As indicated in Appendix B, the graphical construction method used to reduce laboratory data resulted in a variance of predicted TSL. The precise extent of these variances was not determined. The laboratory results presented for each test made are predicted TSL values at various underflow suspended solids concentrations (g/ℓ) and maximum obtainable underflow suspended solids concentration.

Lime-Primary Sludge Thickening

Lime-primary sludge was produced in the chemical treatment unit under a relatively narrow range of operating conditions. Hydraulic loading (0.5 to 1.0 gpm/sq ft), pH (10.8 to 11.2) and pumping turbine speed (24 to 34 rpm) were all closely controlled. The only major operational variable change was constant versus diurnal flow.

Monthly average lime-primary sludge production results ranged from 6100 to 7000 lb of dry solids per MG of plant flow. The weighted average sludge production was 6600 lb/MG. The monthly average lime dosages ranged from 410 to 510 mg/ℓ of $Ca(OH)_2$, with a weighted average of 453 mg/ℓ. This resulted in a weighted average sludge production of 1.7 lb of chemical and primary sludge per lb of lime used (1.66 to 1.86 range). Sludge production values are consistent with those previously reported.[1]

The following assumptions were made in an effort to identify constituents of the lime-primary sludge. These assumptions are based on changes in pertinent water quality parameters caused by lime treatment.

1) Reduction in bicarbonate alkalinity resulted from precipitation of $CaCO_3$,

2) Reduction in non-calcium hardness resulted from precipitation of $Mg(OH)_2$ and

3) Reduction in phoshporus resulted from precipitation of $Ca_5OH(PO_4)_3$.

Based on the above assumptions computed average lime-primary sludge constituents were:

Sewage Solids		13.3
$Mg(OH)_2$		8.3
$Ca_5OH(PO_4)_3$		9.4
$CaCO_3$		69.0
	TOTAL	100.0%

The average computed sludge production value was found to be 95 percent of the observed value of 6600 lb/MG. A calcium balance across the chemical treatment unit indicated some unaccounted calcium precipitate. It is possible that some unreacted lime, $Ca(OH)_2$, and/or calcium-organic compounds were present in the sludge accounting for the slight difference between computed and observed sludge production values. Another probable explanation was that CO_2 was absorbed in stored samples of chemical treatment unit effluent resulting in erroneously high bicarbonate alkalinity determinations on the chemical treatment unit effluent sample. That this may have occurred was indicated by stored composite sample pH's being several tenths of a pH unit lower than daily averaged grab sample pH values. It is, therefore, suggested that the unaccounted for calcium compound in the lime-primary sludge was $CaCO_3$.

Based on the above analysis the lime-primary sludge was about 87 percent chemical precipitates. A very large portion was $CaCO_3$. If the $CaCO_3$ value could be recovered by a classification and recalcination procedure achieving 100 percent recovery of lime, approximately 80 percent of the plant's required lime dosage could be satisfied.

Laboratory test data indicated that the overall lime-primary sludge solids density was 2.0 to 2.6 g/cc. Sludge slurry density was found to be 1.00 plus 0.000553 times the slurry concentration in g/ℓ.

No attempt was made to determine the particle size and distribution of sludge solids. The $CaCO_3$ precipitate formed in municipal wastewater treatment is much more finely divided than that formed in potable water treatment by lime softening. Presumably, the presence of condensed phosphates and soluble organic compounds inhibit particle growth.

Pilot Plant Results

During most of the approximately six months of lime treatment, feed and underflow to and from the thickener was automated. The thickener was fed at 20 to 60 minute intervals at about 200 gpm for 6 to 10 second durations. Thickener blowdown was at the same time interval and at about 150 gpm for 1 to 3 second durations.

During the last 3 weeks of lime treatment, the thickener feed was pumped at a rate of about 5 gpm or less. Also during this period, pickets were attached to the sludge rakes.

Table 7-1 presents average operating and performance data for four periods of reasonably stable thickener operation and performance. The effect of chemical contactor flow variation on underflow solids concentration was quite dramatic. The underflow suspended solids concentration of the chemical contactor was reduced to less than 1/2 at 2:1 peak to minimum, diurnal flow compared to constant flow operation. Constant pumping of the chemical contactor underflow resulted in a slight additional reduction in underflow suspended solids concentrations.

In general, thickener performance at a TSL of 15 to 20 lb/day/sq ft was considered poor. Thickener hydraulic loadings were well below 50 gpd/sq ft resulting in excellent suspended solids capture of 99+ percent. Typical thickener solids profiles shown in Figure 7-1 illustrate the poor thickening performance.

Laboratory Test Results

Twenty-seven laboratory gravity thickening tests were run from November, 1972, to February, 1973.

Table 7-2 presents laboratory test results. The test number denotes the date the sludge sample was grab-composited. For example, #0116 was January 16, 1973, and #1129 was November 29, 1972. Test #'s 0220A, B and C and 0129 were for single grab samples. Sample #0220 was partially thickened and then two portions diluted with supernatant to result in lower suspended solids concentrations.

Numerous attempts were made to define reasonably precise relationships between TSL and feed suspended solids for various desired underflow suspended solids. None were found!

Figure 7-2 shows maximum underflow suspended solids concentrations obtained and the predicted TSL to achieve same for various feed suspended solids concentrations. The linear regression line shown, which has a correlation coefficient of 0.96, does not include data points ⊗, which were obtained for single grab samples of sludge.

The maximum underflow suspended solids concentration is seen to increase directly proportional to the feed suspended solids concentration. This type of response has been reported by others for similar and different types of sludges.[8] It was thought that for a given particle size, distribution and density and chemical character of sludge, that the maximum obtainable underflow suspended solids concentration

TABLE 7-1:

LIME-PRIMARY SLUDGE THICKENING AVERAGE PILOT PLANT RESULTS

Calendar Dates	8/27/72 to 9/27/72	11/8/72 to 12/1/72	1/4/73 to 1/31/73	2/1/73 to 2/22/73
Chemical Contactor:				
Flow:				
Average gpm/sq ft	1.0	1.0	0.85	0.93/0.75
Peak gpm/sq ft	1.0	1.2	0.85	0.93/0.75
Peak:Minimum	1:1	2:1	1:1	1:1
Lime Dosage, mg/ℓ	425	410	510	470
Sludge Production, lb/MG	6600	6100	7000	6700
Turbine Speed, rpm	34	34	26	26
Reaction Zone-SS, g/ℓ	6.8	3.1	2.2	2.8
Bowdown-SS, g/ℓ	177	73	88	67
Gravity Thickener:				
Solids Loading, lb/day/sq ft	20	18	17	~15
Solids Capture, %	99+	99+	99+	99+
Blowdown-SS, g/ℓ	207	109	108	84
Sludge Depth, ft	~5	~5	5	4-5

FIGURE 7-1: LIME PRIMARY SLUDGE THICKENING - THICKENER SOLIDS PROFILE

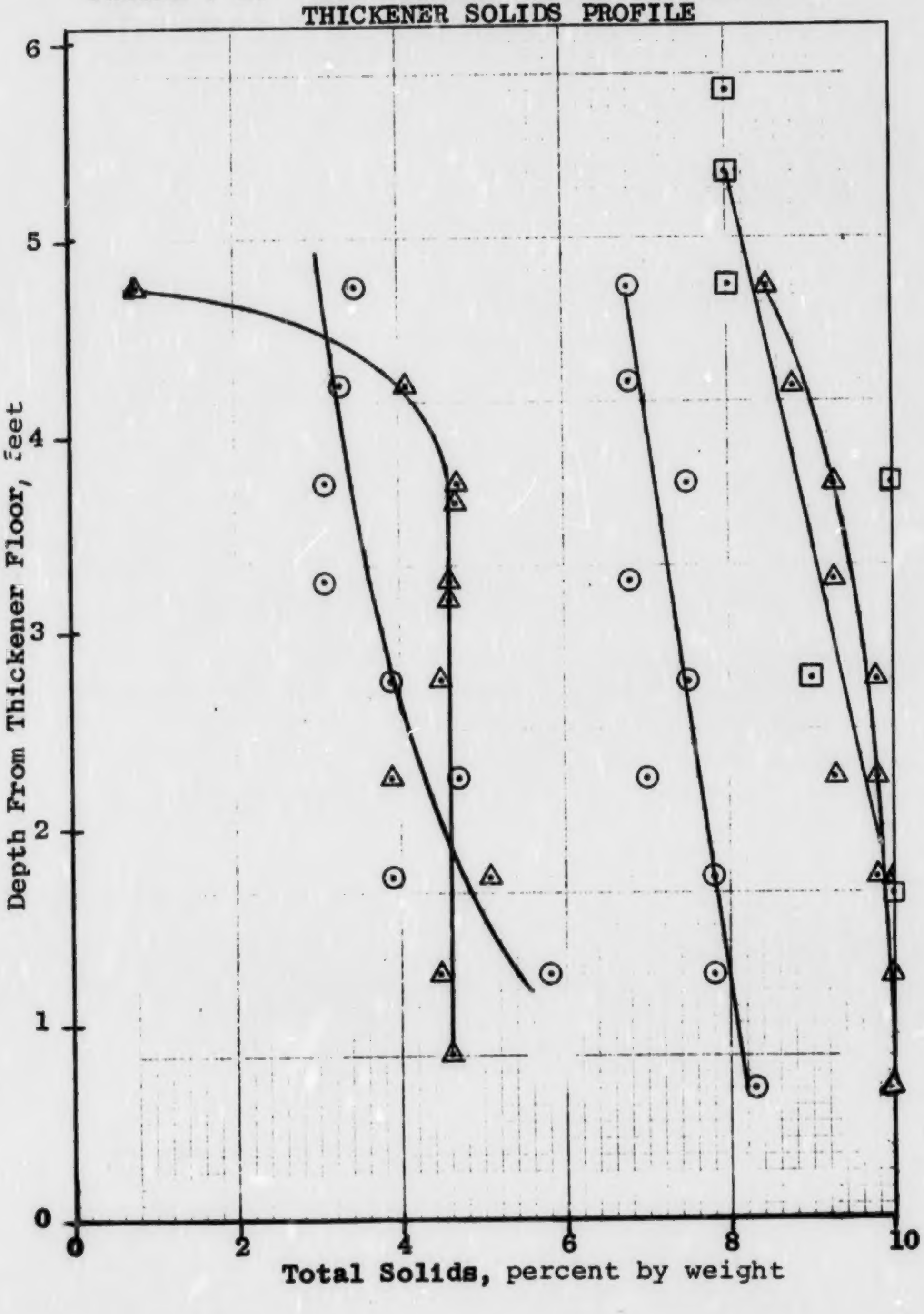
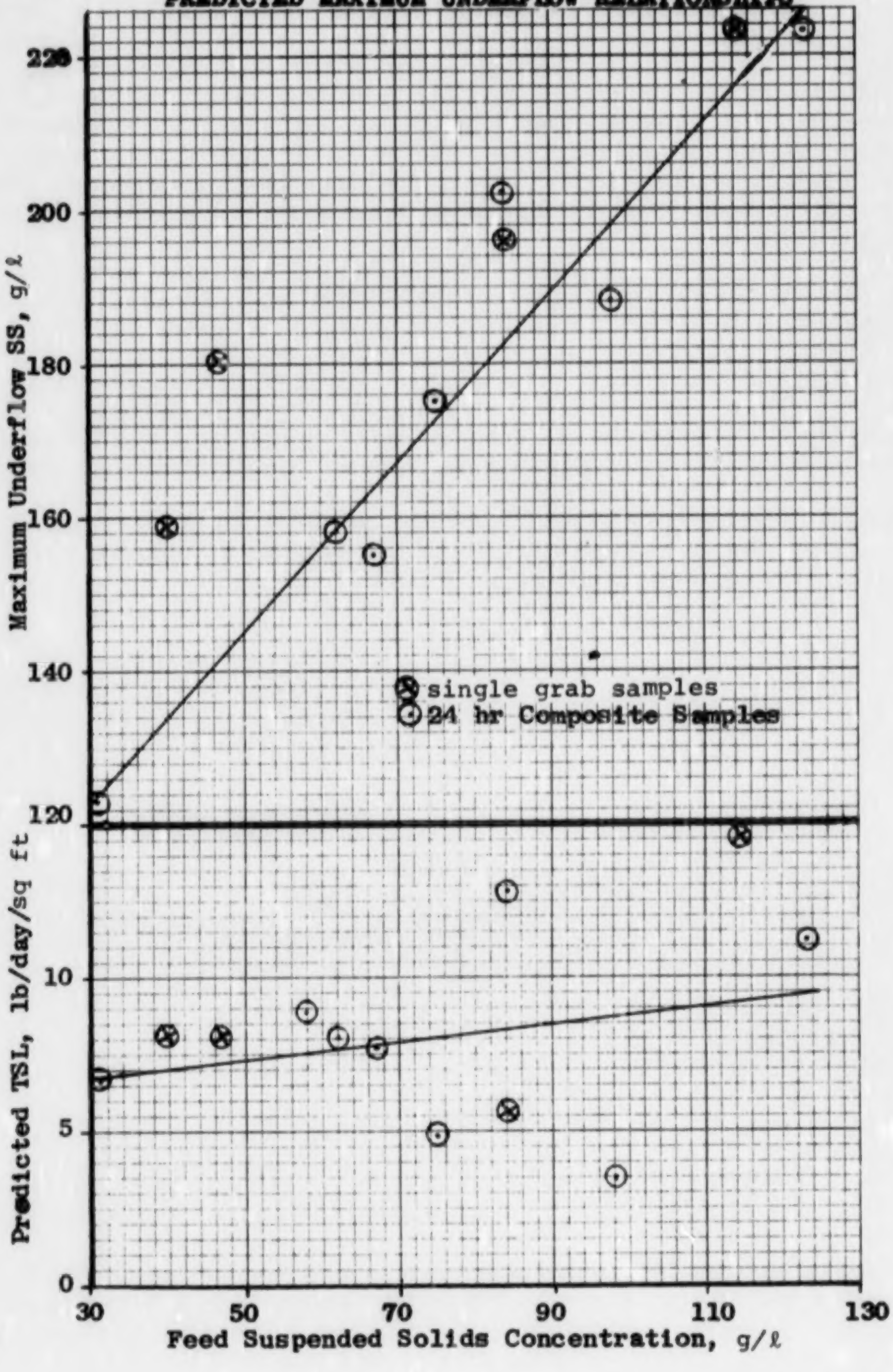


TABLE 7-2: LIME-PRIMARY SLUDGE: LABORATORY THICKENING TEST RESULTS

Test #	Feed SS, g/ℓ	Thickener Solids Loading lb/day/sq ft @ Underflow SS (g/ℓ) of:									Max. Under-flow SS, g/ℓ	Critical Point in inch/7 inch
		80	100	120	140	160	170	180	200	Max.		
0220C[a]	39.5	30	17.1	11.6	9.4	--	--	--	--	8.2	159	1.5
0116	40.4	7.5	5.6	4.8	--	--	--	--	--	--	--	2.1
0220B[a]	46.9	41	24	15.6	11.8	9.9	--	--	--	8.2	180	1.7
0218	56.3	19	10	7.2	6.1	--	--	--	--	--	--	1.1
0206	57.5	71	24	14	11	--	--	--	--	8.9	151	1.5
1129	61.0	35	15	9.6	--	--	--	--	--	--	--	3.6
0204	61.8	48	21	12	9.4	--	--	--	--	8.1	158	2.1
0215	62.9	27	15	9.4	7.3	6.3	--	--	--	--	--	3.4
0114	64.4	86	37	26	21	19	--	18	--	--	--	2.5
0107	65.5	37	13	83	6.6	5.8	--	--	--	--	--	2.4
0213	66.5	46	16	11	8.4	--	--	--	--	7.7	155	2.3
0109	70.4	95	30	16-	12-	9.8	--	8.0	--	--	--	1.7
0125	75.3	--	30	12+	6.4	5.6	--	--	--	4.8	175	1.4
1117	78.0	--	76	34	22	14.5	--	--	--	--	--	2.5
0220	78.4	--	24	18	12	9.5	8.8	--	--	--	--	2.9
0123	84.1	--	106	42	26	18	--	15	--	12.8	202	2.0
0129[a]	84.3	--	62	23	12	8.1	6.7	--	--	5.5	196	1.1
0121	92.3	--	--	29	17	13	--	11	9.7	--	--	2.0
1127	94.5	--	--	73	34	22	19	--	--	--	--	2.8
0130	97.8	--	--	88	38	23	--	9.9	--	3.5	188	1.7
0111	99.2	--	--	44	22	14	12	--	9.9	--	--	2.0
0220A[a]	114	--	--	--	72	36	--	21	17	14.6	223	2.5
0118	123	--	--	--	72	29	--	20	15	11.5	223	1.8

[a] Single grab samples

FIGURE 7-2: LIME PRIMARY SLUDGE THICKENING - LABORATORY PREDICTED MAXIMUM UNDERFLOW RELATIONSHIPS



would be constant. Apparently the thickening properties are, in part, effected by the feed suspended solids concentration. It is probable that the laboratory test procedures used had an effect. As feed suspended solids concentrations increased, the final volume of thickened sludge in the 2 liter graduated cylinder increased slightly. For example, as feed suspended solids increased from 60 to 120 g/ℓ the regression curve on Figure 7-2 indicates that the final volume of thickened sludge increased 45 percent (from 0.38 to 0.55 liters). This deeper sludge layer may have provided increased compressive forces resulting in a higher final suspended solids concentration.

Also shown on Figure 7-2 is the predicted TSL required to achieve the maximum underflow suspended solids concentrations. The linear regression line shown does not have a statistically significant positive slope (at a 95 percent confidence level). This fact indicates that as feed suspended solids concentration increases, higher TSL's are not possible if maximum underflow suspended solids concentration is desired.

A single series of laboratory tests were made to determine the effect of a polyelectrolyte on thickening response. The results of this are shown in Table 7-3.

TABLE 7-3:

LIME-PRIMARY SLUDE THICKENING:
EFFECT OF POLYELECTROLYTE

Test No.	Feed SS g/ℓ	Thickener Solids Loading (lb/day/sq ft) @ Underflow SS (g/ℓ) of:					Max. Under-Flow SS, g/ℓ	Critical Point inches/ 7 inch	Polymer Dosage lb/Ton Dry Solids
		50	60	80	100	120			
222-1	31	22	13	8.3	6.8	--	>>89	1.2	0.6
222-2	31	--	42	19	13	11	>116	1.1	1.0
222-3	31	--	65	29	18	14	123	1.4	1.9
222-4	31	--	--	61	48	42	>116	2.6	3.9

The polyelectrolyte used was anionic AltaSep 2A2. No screening of other potentially effective polyelectrolytes was made. Previous studies had indicated that the Altasep 2A2 was very effective in increasing the settling rate of relative dilute slurries (about 800 mg/ℓ) resulting from lime treatment.[1] The dramatic effect of polyelectrolyte is shown on Figure 7-3. It should be noted, however, that the feed suspended solids concentration was quite low (31 g/ℓ). The effect of polyelectrolyte on predicted TSL for higher feed suspended solids would be expected to be less dramatic. As seen in Table 7-3, the maximum underflow suspended solids concentration reached did not seem to be greatly effected by use of polyelectrolyte. The major effect is seen to be a dramatic increase in initial sludge subsidence rate as shown on Figure 7-4.

Comparison of Pilot Plant and Laboratory Results

It was originally intended to collect laboratory and pilot plant data so that a direct comparison could be made. The intended approach was to, before the fact, identify days of reasonably stable pilot plant thickener operation. Performance data would be collected and a grab-composite sample obtained for laboratory tests. Because of an inability to predict operational problems before the fact, judgement was not always sound. Of the 19 laboratory tests conducted for the above purpose, only five comparisons indicated similar laboratory and pilot plant results. The lack of precise correlation between pilot plant and laboratory results presumably was due, in part, to variability of plant operating conditions, sludge sampling, laboratory test procedures and/or laboratory data reduction procedures.

In general, the laboratory predicted TSL and underflow suspended solids concentrations were considerably in excess of pilot plant results. For example, the ratio of underflow suspended solids concentrations achieved by the pilot plant to that predicted by the laboratory results, at the actual pilot plant TSL, ranged from 0.50 to 1.09, with an average of 0.75 for the 19 comparisons made. The ratio of pilot plant TSL to laboratory predicted TSL, at the pilot plant underflow suspended solids experienced, was much more variable.

The lack of a definable precise correlation based on one on one comparisons of pilot plant and laboratory results prompted a different approach. Inspection of pilot plant performance data in Table 7-1 shows that the average TSL was just under 18 lb/day/sq ft. The laboratory predicted underflow suspended solids concentrations at a TSL of 18 lb/day/sq ft were determined, from plots of data in Table 7-2, and then plotted vs feed suspended solids as per the circles on Figure 7-5. The linear regression curve shown for these

FIGURE 7-3:

LIME PRIMARY SLUDGE THICKENING -
EFFECT OF POLYELECTROLYTE ON PREDICTED TSL

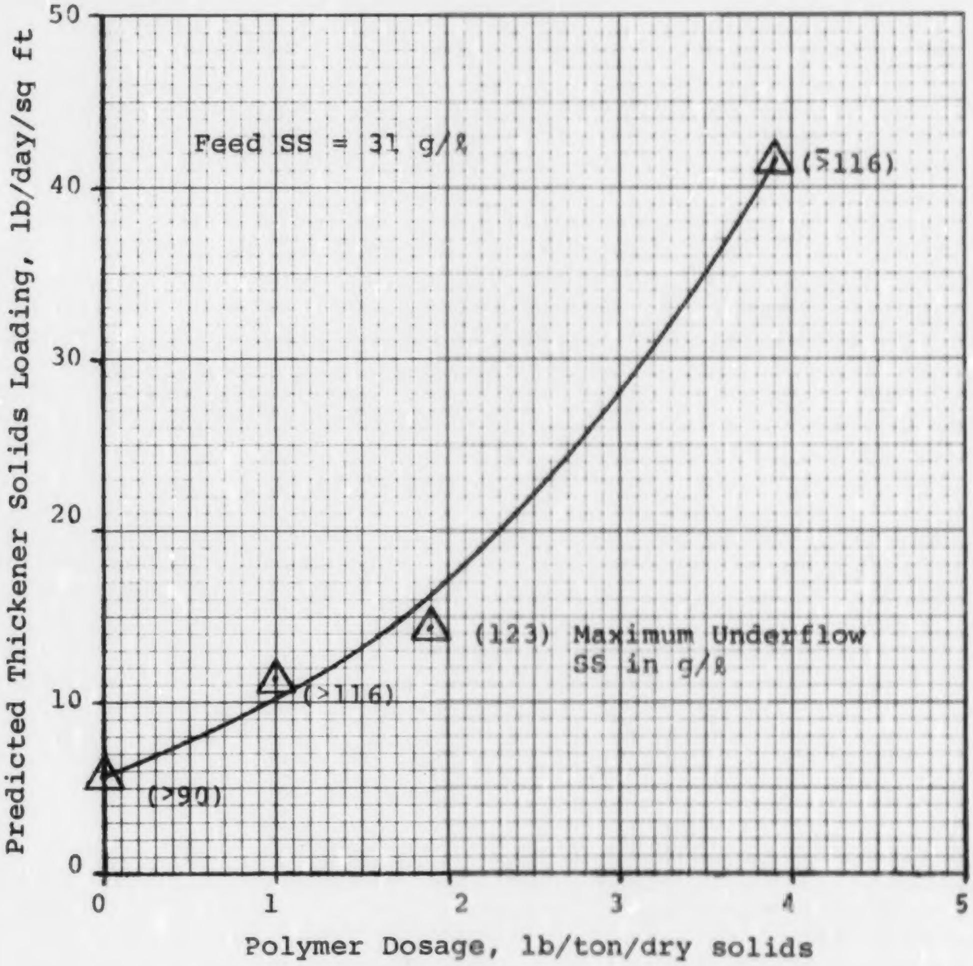


FIGURE 7-4: LIME PRIMARY SLUDGE THICKENING - EFFECT OF POLYELECTROLYTE ON SETTLING RATE

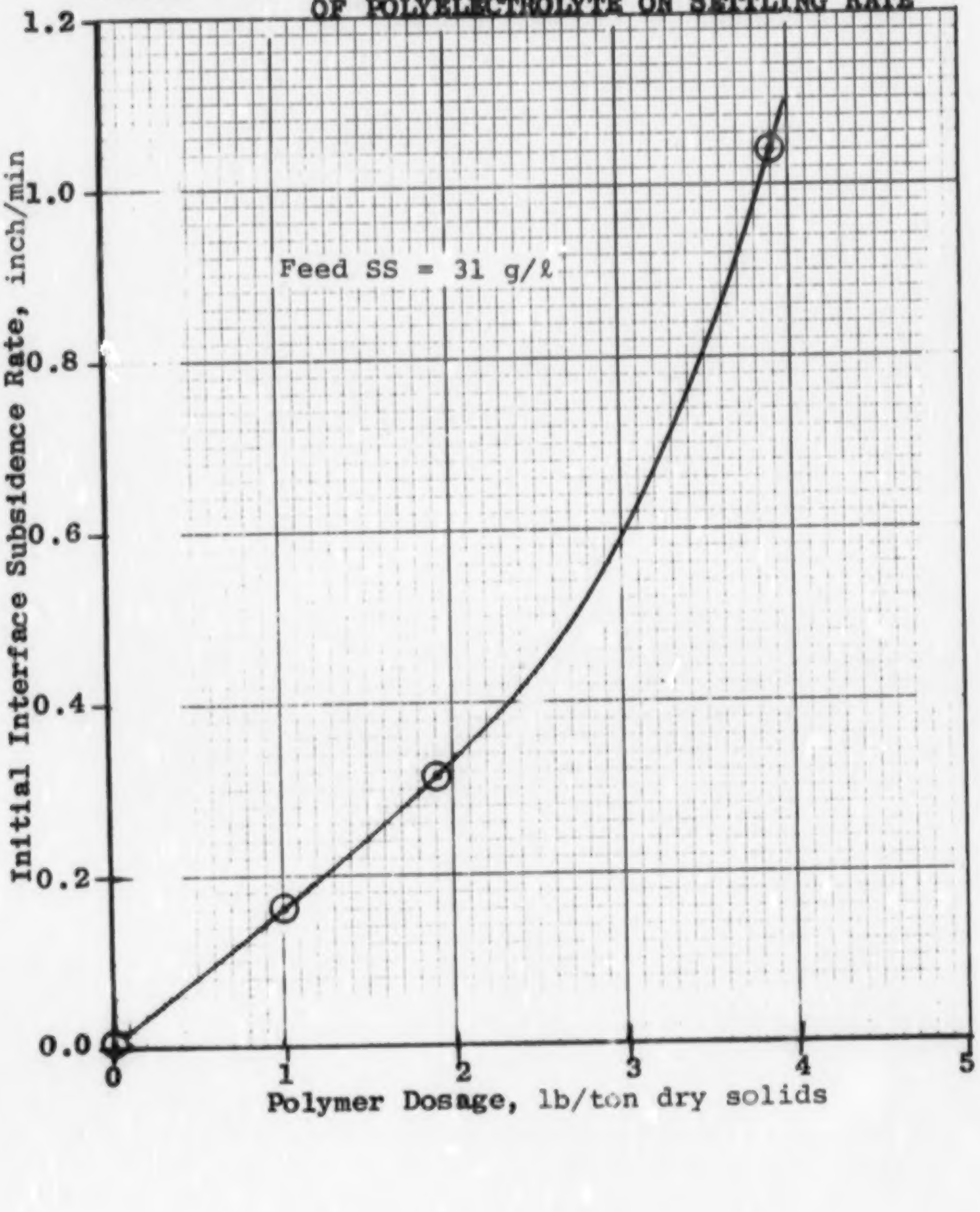
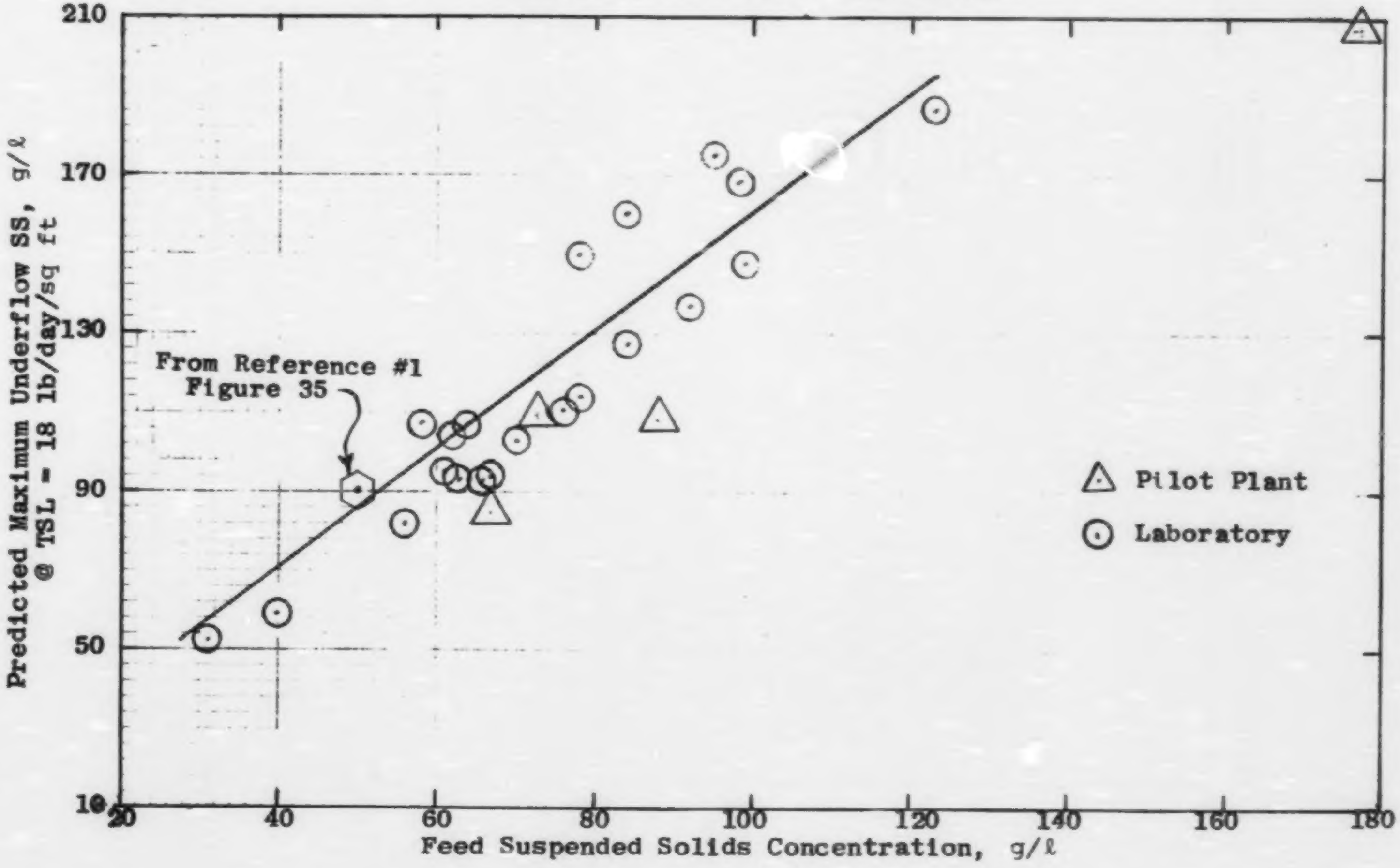


FIGURE 7-5: LIME PRIMARY SLUDGE THICKENING - COMPARISON OF LABORATORY AND PILOT PLANT RESULTS



data has a correlation coefficient of 0.91. The triangles on Figure 7-5 are pilot plant results from Table 7-1. It is readily apparent that laboratory predicted values are generally in excess of pilot plant performance. The limited number of pilot plant data points on Figure 7-5 does not allow statistical definition of a scale-up factor.

For comparative purposes the ⊙ data point on Figure 7-5 is shown. This data point was obtained from Figure 35 of the previous contract report for laboratory test results.[1] The closeness of fit with current contract results implies a similarity in the two sets of data.

Summary Discussion

As previously noted pilot plant performance was less than desired and expected. Initially it was felt that the current results could not be explained, based on previous contract experience. After evaluation of the data it becomes apparent that the reason for less than desired thickener underflow suspended solids concentration was the reduced chemical clarifier underflow suspended solids concentration experienced due to diurnal flow and poor plant operation. This observation indicates the need to precisely determine the clarifier underflow to be experienced prior to designing prototype thickeners. It may be more rational to use a significant safety factor in predicting clarifier underflow suspended solids concentration (i.e., thickener feed), than to use a safety factor in predicting thickener performance.

A precise correlation between pilot plant and laboratory test thickening results was not defined. The ratios of pilot plant to laboratory test predicted underflow suspended solids concentrations shown on Figure 7-5 were 0.75, 0.91 and 0.76 for the three data points shown. These data were for similar pilot plant and laboratory test TSL values. This observation suggests an approach to using the laboratory test data for predicting pilot plant results. The following is an example of this suggested approach.

Assume a thickener feed suspended solids concentration of 80 g/ℓ. If maximum thickener underflow is desired, Figure 7-2 indicates that a TSL of 8 lb/day/sq ft should be used. The laboratory data predicted the underflow would be 178 g/ℓ. If the pilot plant to laboratory predicted ratio existing for data on Figure 7-5 were applicable, a scale-up factor of about 0.75 would apply. Thus the expected underflow would be 0.75 x 178 or 134 g/ℓ.

A thickener operating with a 3 ft deep sludge layer, a 134 g/ℓ underflow suspended solids and a TSL of 8 lb/day/sq ft would have a sludge SRT of less than 3 days,

Alum-Primary Sludge Thickening

Alum-primary sludge was produced and thickened for just over a 3 month period during April, May and October of 1973. Analysis of thickened sludge inventory results and chemical clarifier feed and effluent suspended solids indicated that the sludge contained approximately 15 to 20 percent, by weight, chemical solids. Slurry densities ranged from 1.05 g/cc for about 20 g/ℓ slurries to 1.10 g/cc for 60 g/ℓ slurries.

Pilot Plant Results

The pilot plant was operated with diurnal flow variation during the entire alum treatment period. The chemical clarifier blowdown was at a rate of about 200 gpm for 6 to 10 seconds at 60 ± 20 minute intervals. The blowdown flowed by gravity to a 150 gal surge tank from which the sludge was pumped to the thickener at less than 5 gpm. During April and May, 1973, thickener blowdown was automated. The weighted average duration of blowdown was 4.5 seconds (range 3 to 10) and the weighted average frequency of blowdown was each 75 minutes. During October sludge was manually blowndown from one to five times each day by plant operators in order to maintain the sludge level between two specified sample taps. Thickener blowdown rate was about 150 gpm for both automatic and manual operation.

Table 7-4 presents average results for three periods of reasonably stable thickener operation and performance. The effect of going from a 2:1 to 3:1 peak:minimum diurnal flow variation on chemical clarifier underflow suspended solids concentration was substantial. The slightly higher average hydraulic loading was probably partly responsible for the reduction in underflow suspended solids from 27 to 16 g/ℓ. Comparison of April and October, 1973, results indicate that the use of 0.25 mg/ℓ of anionic polyelectrolyte in the chemical unit did not result in expected higher underflow suspended solids concentration.

It is interesting to note that the underflow suspended solids concentrations of 15 to 27 g/ℓ achieved were from 4 to 15 times greater than experienced during previous contract work at constant flow conditions. Obviously the baffle plate shown on Figure 4-3 was very effective in isolating the sludge withdrawal region from the mixing reaction region.

TABLE 7-4

ALUM-PRIMARY SLUDGE THICKENING AVERAGE PILOT PLANT RESULTS

Calendar Date	4/10/73 to 4/26/73	5/13/73 to 5/21/73	10/17/73 to 11/9/73
Chemical Contactor:			
Flow:			
Average gpm/sq ft	0.35	0.43	0.43±0.10
Peak gpm/sq ft	0.45	0.60	0.55±10
Peak:Minimum	2:1	3:1	2:1
Alum Dosage, mg/ℓ	140	120	120
Polymer Dosage, mg/ℓ	0	0	0.25
Sludge Production, lb/MG	1700	1100	1200
Turbine Speed, rpm	21	21	21→18
Reaction Zone-SS, g/ℓ	1.5	1.7	0.93
Blowdown-SS, g/ℓ	27	15.6	15.3
Gravity Thickener:			
Solids Loading, lb/day/sq ft	1.7	7.1	7.8
Sludge Level, ft	3→1	1-2	2-4
Solids Capture, %	99	99	99
Blowdown-SS, g/ℓ	75	28	28
	7-3	1-2	2-4

As the thickener data in Table 7-4 show, during April the TSL was a very low 1.7 lb/day/sq ft. Thickener underflow suspended solids were an impressive 75 g/ℓ. Of importance, however, is the relative long sludge SRT of 3-7 days. During start-up of the thickener, about 2 weeks was required to buildup a sludge depth of 4 ft. At that time, thickener blowdown was started and the sludge level gradually reduced to a 1 ft depth over a 2-1/2 week period. After 1-1/2 weeks, the sludge turned septic and by the end of the 2-1/2 week period was black, very odorous and releasing minute gas bubbles.

On April 26, the 8 ft diameter thickener was taken off stream. A 3.5 ft diameter by 6 ft deep unit was fabricated, installed inside the 8 ft unit and brought on stream in early May. No septicity problems were experienced during subsequent operation at approximate sludge SRT's of 1 to 4 days during May and October, 1973.

Data in Table 7-4 indicates that the use of polyelectrolyte in the chemical unit did not improve pilot plant thickener performance. This observation was contrary to what was expected.

Figure 7-6 presents several thickener solids vs depth profiles determined during the period polyelectrolyte was used. It is seen that about 2 to 3 ft of sludge was required to reach a maximum thickened solids concentration.

Laboratory Test Results

Laboratory results are presented in Tables 7-5 and 7-6. All tests shown were run on 24 hour grab-composite samples. Inspection of the "critical point" data indicates an average of 1.2 for Table 7-5 and 2.2 for Table 7-6. Consequently, predicted TSL in Table 7-5 are relatively more conservative than in Table 7-6 (see Appendix B for discussion of critical point).

As indicated for the lime-primary sludge results, there is a general increase in maximum obtainable underflow suspended solids concentrations as feed suspended solids increases. This data, plotted on Figure 7-7, shows that no precise relationship exists for all combined data. There is a definite trend in the April and May results (series 400 and 500 test #'s). Correlation coefficients for the two sets of data shown was found to be 0.89 and 0.95, respectively. The data obtained when polyelectrolyte were used in the chemical unit (Table 7-6 and ⊡'s on Figure 7-7) are not well correlated. For this data, the average feed suspended

FIGURE 7-6: ALUM PRIMARY SLUDGE THICKENING - THICKENER SOLIDS PROFILE

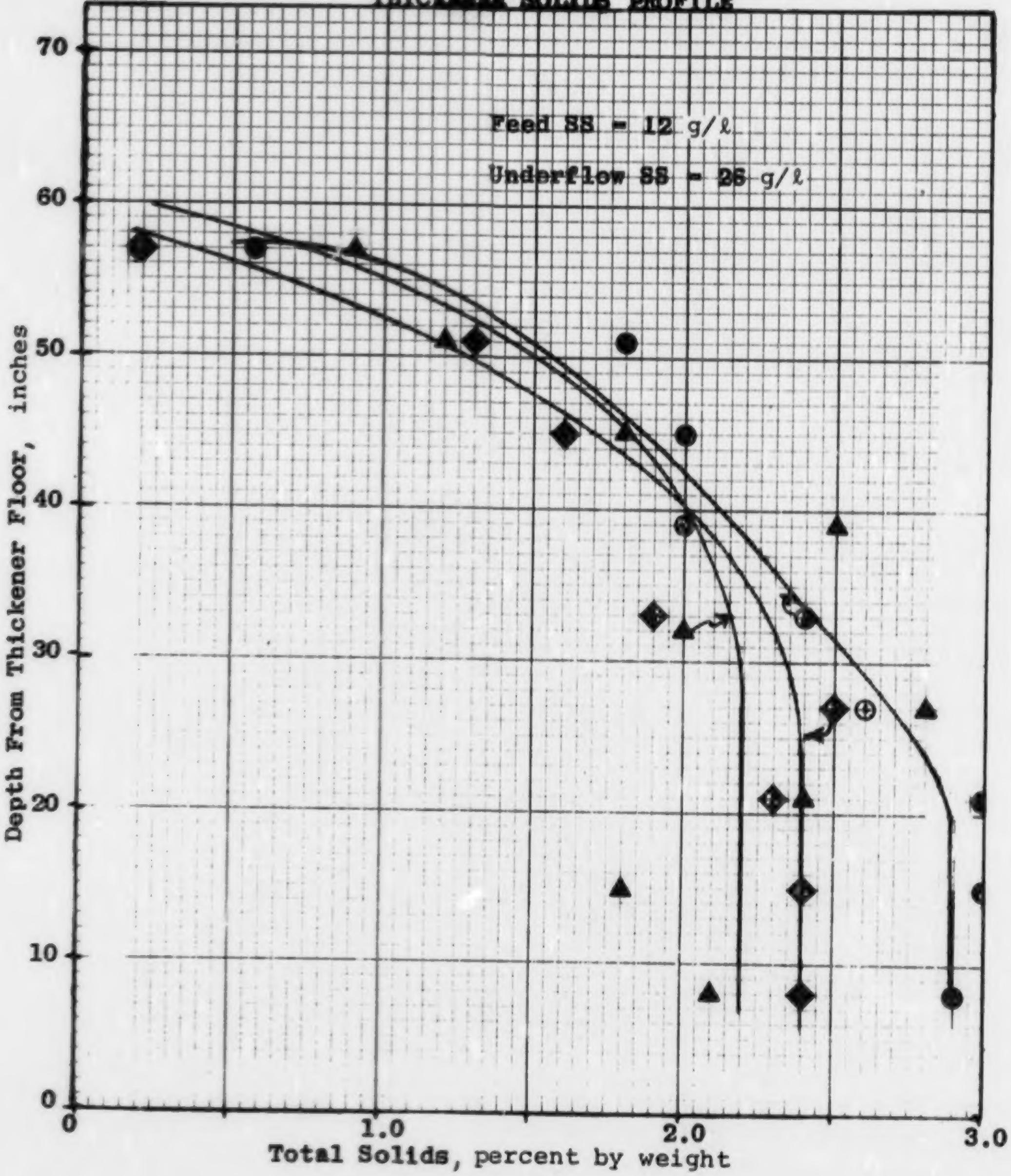


TABLE 7-5: ALUM-PRIMARY SLUDGE: LABORATORY THICKENING TEST RESULTS[a]

Test #	Feed SS, g/1	Thickener Solids Loading, lb/day/sq ft @ Underflow SS (g/1) of:							Max. Under-flow SS,g/1	Critical Point in Inch/7 Inch
		20	30	40	50	60	70	Max.		
512	11.5	27	6.9	5.3	--	--	--	5.3	43	0.9
506	13.9	42	24.0	15.0	7.1	--	--	5.0	60	0.9
410	15.2	--	13.0	8.1	4.5	--	--	4.0	55	1.0
412	16.1	--	21.0	9.9	5.13	--	--	5.3	49	0.9
510	17.0	67	24.0	11.0	5.4	--	--	4.0	57	1.2
429	17.7	107	31.0	16.4	6.7	--	--	5.4	55	1.0
426	19.4	--	38.0	23.7	12.7	--	--	10.3	54	1.0
405	20.5	--	--	23.0	13.0	9.0	--	7.4	70	1.8
501	22.1	--	58.0	35.0	20.0	9.8	--	7.4	68	0.9
422	24.3	--	52.0	33.0	19.0	11.8	--	10.8	64	1.8
424	25.2	--	109.0	47.0	24.0	14.0	10.9	11.5	67	1.5
520	29.5	--	--	78.0	52.0	43.0	32.0	~12	98	1.0
417	32.7	--	--	74.0	37.0	19.0	14.0	13.8	71	1.6
419	41.7	--	--	--	56.0	28.0	16.0	9.2	82	1.7
325	50.2	--	--	--	--	35.0	18.0	9.5	81	2.1

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[a] without polyelectrolyte to chemical contactor

TABLE 7-6: ALUM-PRIMARY SLUDGE: LABORATORY THICKENING TEST RESULTS[a]

Test #	Feed SS, g/1	Thickener Solids Loading, lb/day/sq ft @ Underflow SS (g/l) of:							Max. Under-flow SS,g/1	Criti-cal Point Inch/7 Inch
		20	25	30	35	40	50	Max.		
1101	10.1	22	15	10.3	--	--	--	8.2	33	2.1
1030	12.4	36	--	12.0	9.8	8.4	--	8.2	42	2.3
1023	12.6	73	--	23.0	--	12.6	9.9	9.9	51	1.8
1106	14.9	75	35	20.0	13.8	--	--	13.3	36	2.2
1028	15.8	64	28	17.0	--	10.1	--	10.7	39	2.3
1104	22.3	--	--	60.0	--	24.0	15.0	15.0	50	2.3
1009	24.9	--	--	67.0	--	21.0	12.6	13.5	48	2.5

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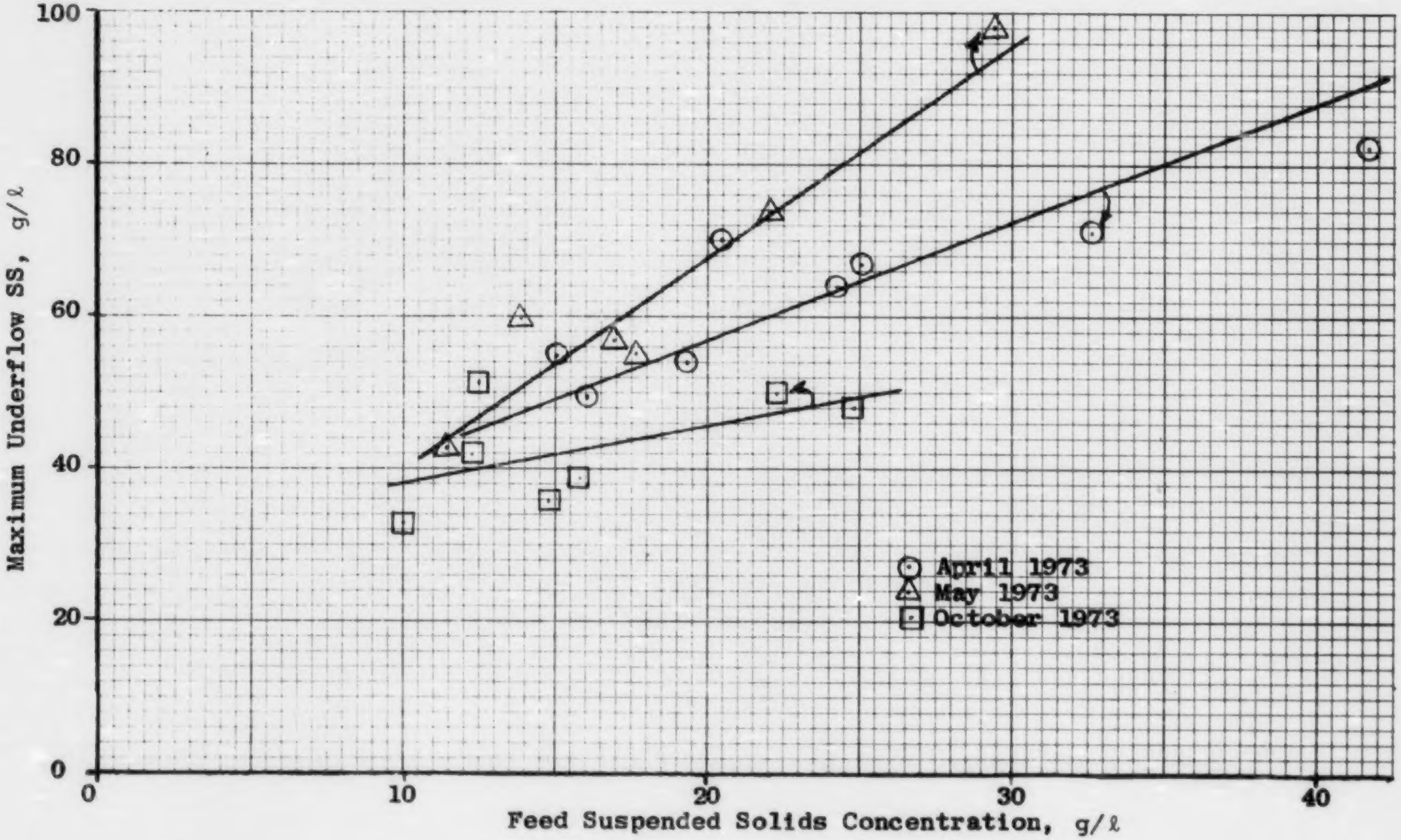
[a] with polyelectrolyte to the chemical contactor

FIGURE 7-7: ALUM PRIMARY SLUDGE THICKENING - MAXIMUM LABORATORY UNDERFLOW CONCENTRATION



solids was 15.3 g/ℓ and the average maximum underflow suspended solids was 43 g/ℓ.

Interrelationships between predicted TSL, feed suspended solids and underflow suspended solids are shown on Figures 7-8 and 7-9. It is apparent that relatively precise relationships were found. Comparison of predicted TSL curves for underflow suspended solids of 40 g/ℓ and the maximum g/ℓ on Figure 7-8 and 7-9, for the feed suspended solids in the 15 to 25 g/ℓ range, indicates the use of polyelectrolyte in the chemical clarifiers resulted in substantially higher TSL's. The difference diminished to a very small value at 25 g/ℓ. It should be noted, however, that in the 15 to 25 g/ℓ feed suspended solids range, maximum underflow suspended solids during April, without polyelectrolyte, were considerably higher than during October with polyelectrolyte (~60 vs 43 g/ℓ).

Comparison of Pilot Plant and Laboratory Results

Eighteen one on one comparisons between pilot plant and laboratory test results were made. Of these, fourteen laboratory test results predicted that the pilot plant should have achieved maximum obtainable underflow suspended solids. This observation is realistic for the April data when the pilot plant thickener was grossly underloaded. That this condition existed is indicated by the ☆ data point on Figure 7-8, which represents average pilot plant data from Table 7-4. During the operating period indicated, the ratio of pilot plant to laboratory underflow suspended solids concentrations was 75/65, or 1.2. From Figure 7-8 it is seen that the laboratory predicted TSL to reach maximum underflow suspended solids was just over 10 lb/day/sq ft, which was about 6 times the average pilot plant TSL. It was this observation that prompted installation of a new pilot plant thickener of about one-fifth the area of the original 8 ft diameter unit.

Only one laboratory test was run during the May operating period indicated in Table 7-4. Thus a direct comparison is impossible. The April laboratory results on Figure 7-8 indicates that a TSL of 7.1 (the May data average) an underflow suspended solids of just over 40 g/ℓ should have been achieved. The 28 g/ℓ actually achieved is 70 percent of the 40 g/ℓ.

Seven laboratory tests during October, 1973, indicated that, at the average TSL of 7.8 (see Table 7-4), an average maximum underflow of 43 g/ℓ should have been achieved. The 28 g/ℓ actually achieved is 65 percent of the predicted

FIGURE 7-8: ALUM PRIMARY SLUDGE THICKENING - PREDICTED TSL VERSUS FEED SUSPENDED SOLIDS CONCENTRATION AT VARIOUS UNDERFLOW CONCENTRATIONS

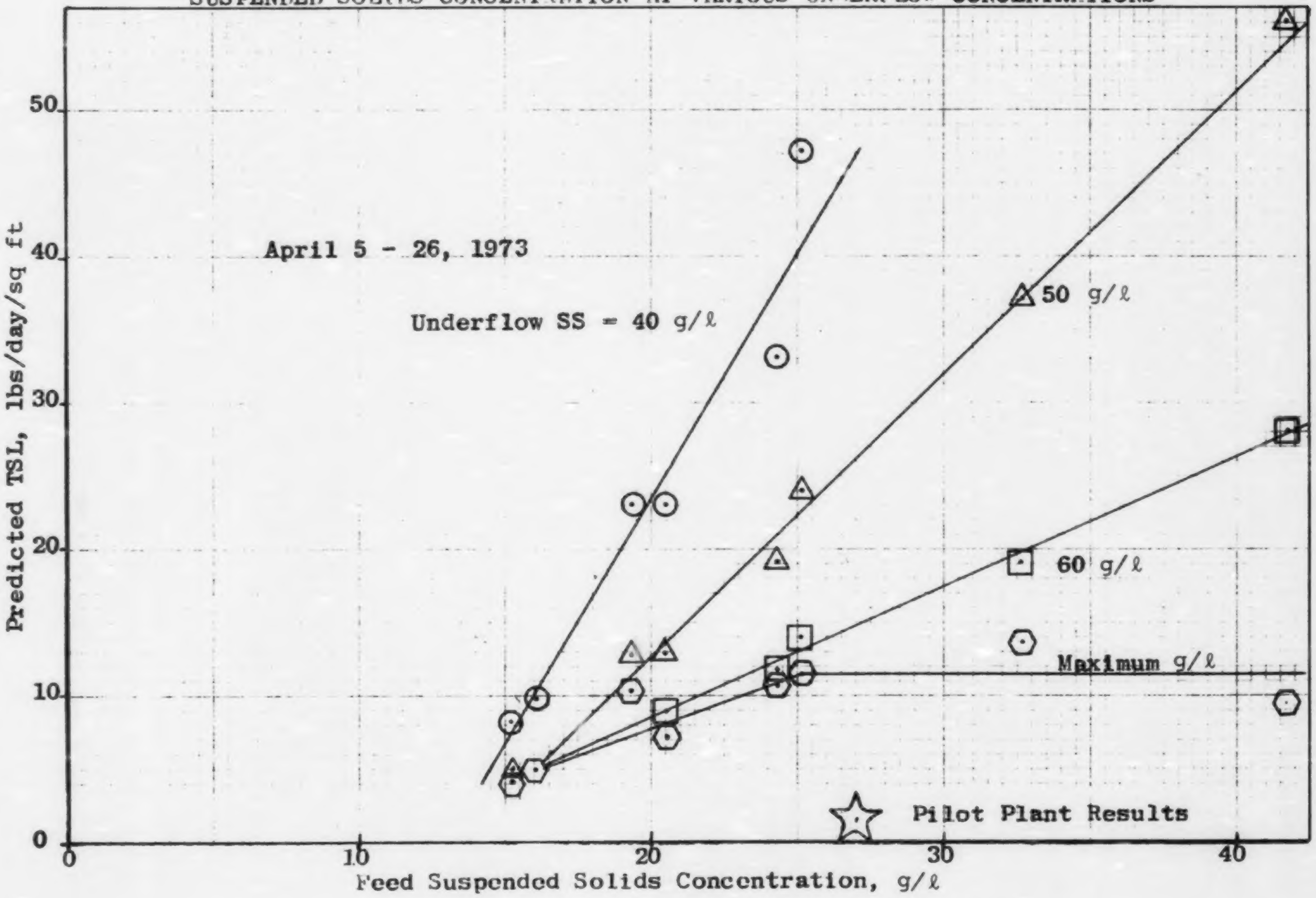
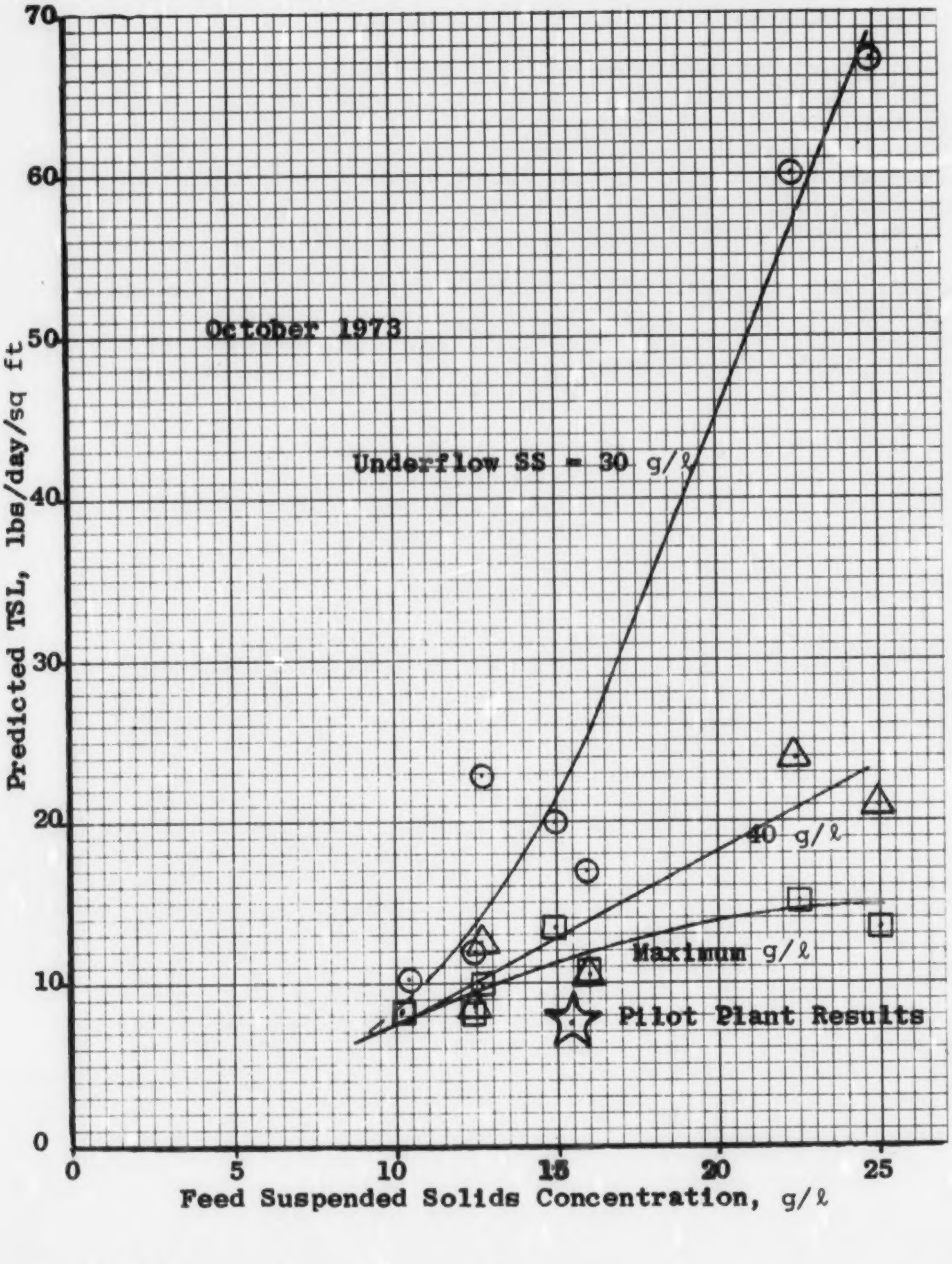


FIGURE 7-9 : ALUM PRIMARY SLUDGE THICKENING - PREDICTED TSL VERSUS FEED SUSPENDED SOLIDS CONCENTRATION AT VARIOUS UNDERFLOW CONCENTRATIONS



43 g/ℓ. The average pilot plant data for the October period is shown on Figure 7-9 as a ☆. It can be seen that to produce an underflow of 28 g/ℓ, the laboratory data predicted a TSL of just over 23 lb/day/sq ft. The actual TSL of 7.8 lb/day/sq ft is 33 percent of the predicted 23 lb/day/sq ft.

Summary Discussion

Pilot plant results indicated that sludge SRT's in excess of 3 to 4 days should be avoided to prevent septic sludge conditions. Pilot plant results also indicated that use of polyelectrolyte in the chemical unit did not increase pilot plant chemical unit or thickener underflow suspended solids concentration. This result was contrary to expectations and to laboratory test results.

Comparison of pilot plant and laboratory test results were similar to that for the lime-primary sludge data. That is to say, for an underloaded thickener, at a given TSL the pilot plant underflow suspended solids concentration was 0.65 to 0.70 of that predicted by laboratory test results. This observation implies that had the pilot plant thickener been operated at a TSL of 10 to 12 rather than the 7.8 lb/day/sq ft shown in Table 7-4, that the underflow suspended solids concentration would have been the same. The validity of this conjecture is doubtful.

Ferric-Primary Sludge Thickening

The pilot plant was operated for just over three months using ferric chloride in the chemical treatment unit. Analysis of thickened sludge inventory data and chemical treatment unit feed and effluent suspended solids concentrations indicated that the ferric-primary sludge was about 15 to 30 percent chemical solids. It was conjectured in Section VI that considerable ferric iron was reduced to soluble ferrous iron within the chemical treatment unit. After a few weeks of operation, the chemical clarifier underflow sludge turned black due to the presence of a finely divided black precipitate. Qualitative analysis of the sludge solids indicated the presence of considerable sulfides which were released as H_2S upon acidification. Quantitative analysis for soluble sulfides indicated only trace amounts. It was concluded that considerable FeS was present in the ferric-primary sludge. The exact amount was not determined. Though the sludge was black, it had no other characteristics of septicity (e.g., odors or gas bubbles).

Pilot Plant Results

Average pilot plant operating and performance data are presented in Table 7-7. It is seen that the only operational variation was the use of a reduced ferric chloride dosage and 0.25 mg/ℓ of anionic polyelectrolyte during the last month of operation. Contrary to results with alum treatment, the use of polyelectrolyte produced a substantial improvement in the chemical unit underflow suspended solids concentration (from 11 to 23 g/ℓ).

The only thickener operational difference between the first and second time periods shown in Table 7-7 was the method of thickener blowdown. During the first period blowdown was automated. At about 20 to 40 minutes intervals approximately 150 gpm was blowndown for 1 to 5 second durations. Because of this automated feature, minimal operator attention was given to the thickener. Consequently, wide variations in sludge depth were experienced. In addition, frequent clogging of the thickener feed pump occurred. This caused chemical clarifier blowdown to overflow the surge tank directly into the 430 gal thickener at about 200 gpm for 6 to 10 seconds each 20 to 40 minutes. The result was a high level of turbulence in the thickener. The two very steep total solids vs depth profiles shown on Figure 7-10 were typical during this operating period. Thickener underflow suspended solids ranged from 8 to 20 g/ℓ and averaged only 15 g/ℓ.

During the second operating period in Table 7-7, thickener blowdown was manually accomplished. Sludge depth was maintained at a specified level, plus or minus 1 ft. In addition, the thickener feed pump operation was frequently monitored, minimizing hydraulic surges to the thickener. Thickener blowdown suspended solids data in Table 7-7 show that more careful operation of the thickener resulted in a doubling of concentration from 15 to 31 g/ℓ. A slight improvement in solids capture was also experienced.

During the last operating period, when polyelectrolyte was added to the chemical unit, thickener performance was dramatically improved. This improvement is shown by results in Table 7-7 and sludge solids profiles on Figures 7-10.

Laboratory Test Results

The results of twenty laboratory tests are presented in Tables 7-8 and 7-9. These tests were conducted on samples with a very limited range of feed suspended solids concentrations. Consequently, any interrelationship between TSL and underflow suspended solids and feed suspended solids

TABLE 7-7

FERRIC-PRIMARY SLUDGE THICKENING AVERAGE PILOT PLANT RESULTS

Calendar Dates	6/10/73 to 7/20/73	7/21/73 to 8/8/73	8/27/73 to 9/17/73
Chemical Contactor:			
Flow:			
Average gpm/sq ft	0.38	0.38	0.39
Peak gpm/sq ft	0.50	0.50	0.50
Peak:Minimum	2:1	2:1	2:1
$FeCl_3$ Dosage, mg/ℓ	98	98	62
Polymer Dosage, mg/ℓ	0	0	0.25
Sludge Production, lb/MG	1100	1200	1000
Turbine Speed, rpm	12-16	12-16	23
Reaction Zone-SS, g/ℓ	1.0	1.0	1.3
Blowdown-SS, g/ℓ	11	11.2	24
Gravity Thickener:			
Solids Loading,	6.3	6.7	5.8
lb/day/sq ft	erratic	~ 4	2-4
Sludge Depth, ft	95+	97+	99
Solids Capture, %	15	31	56
Blowdown-SS, g/ℓ			

FIGURE 7-10: FERRIC PRIMARY SLUDGE THICKENING - THICKENER SOLIDS PROFILE

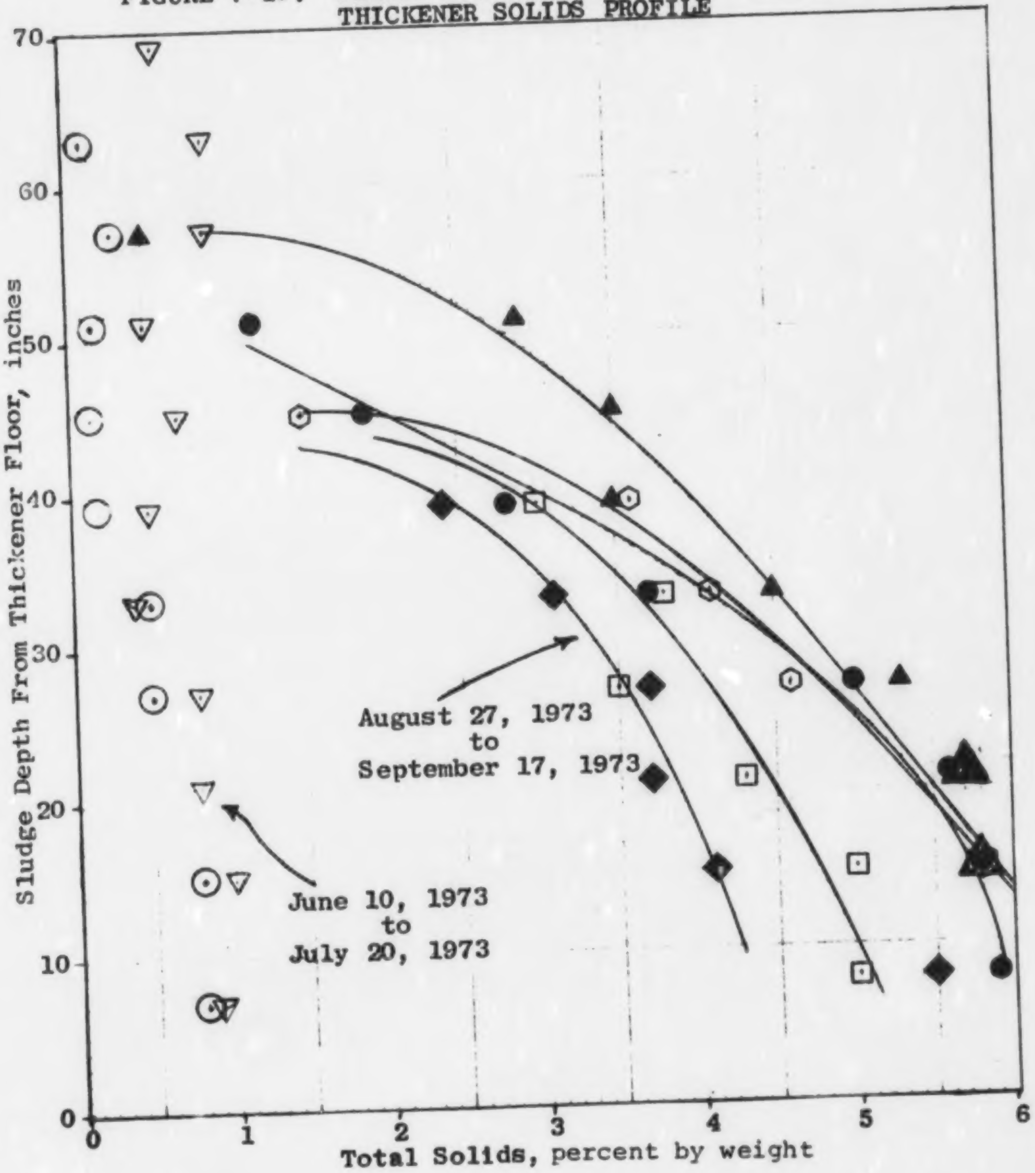


TABLE 7-8: FERRIC CHLORIDE-PRIMARY SLUDGE: LABORATORY GRAVITY THICKENING TEST RESULTS

Test #	Feed SS, g/1	Thickener Solids Loading, lb/day/sq ft @ Underflow SS (g/l) of:						Max. Under-flow SS,g/1	Criti-cal Point in Inch/7 Inch
		10	20	30	40	50	Max.		
621	2.2	9.2	4.5	3.9	3.7	--	3.6	43	1.3
705	5.7	28	14	7.2	5.7	5.1	4.6	57	1.6
619	6.1	12	3.4	1.7	--	--	1.5	36	1.3
805	7.0	36	18	11	8.1	6.8	6.3	61	1.4
612	7.3	21	7.7	3.0	2.1	--	1.8	48	1.2
626	7.3	23	6.4	2.4	1.8	1.6	1.6	52	1.3
617	8.0	16	3.4	1.7	--	--	1.6	33	1.9
701	8.0	--	9.0	5.2	4.3	3.9	3.7	57	2.3
624	8.2	65	24	10	4.8	3.4	2.6	68	1.1
610	9.0	--	8.8	3.8	2.5	2.1	2.0	57	1.2
726	9.4	--	--	16	10	8.1	6.5	75	1.9
708	10.9	--	49	29	17	13	11.0	58	1.3
703	11.0	--	39	23	14	8.0	4.6	71	0.9
Averages:	7.7						4.0	55	1.4^+

TABLE 7-9:

FERRIC CHLORIDE-PRIMARY SLUDGE: LABORATORY GRAVITY THICKENING TEST RESULTS

Test No.	Feed SS g/ℓ	Thickener Solids Loading (lb/day/sq ft) @ Underflow SS (g/ℓ) of: 30	40	50	Max	Max. Underflow SS, g/ℓ	Critical Point inch/7 inches
907	20.8	89	45	28	23	60	2.7
911	21.3	95	40	22	18	57	1.8
830	21.7	94	38	20	18	52	2.2
913	25.5	123	31	--	19	47	2.1
910	26.1	202	53	22	20	53	2.1
Averages	23.1				20	54	2.2

concentrations cannot be defined. For this reason, average results are shown in Tables 7-8 and 7-9.

Comparison of average maximum obtainable suspended solids concentrations from Tables 7-8 and 7-9 indicate no difference. The higher feed suspended solids concentrations and polymer dosage for Table 7-9 results apparently had no effect on maximum suspended solids concentration. Recall that for both lime and alum-primary sludges, increased feed suspended solids results in increased underflow solids concentrations. The average predicted TSL's to reach maximum underflow solids when using polyelectrolyte in the chemical unit is seen to be five times higher than when polyelectrolyte was not used. This direction of change was expected.

Evaluation of data in Table 7-8 indicates a general positive trend of increasing maximum underflow solids concentrations and predicted TSL at those concentrations as feed suspended solids increases. The data ranged from about 1.5 lb/day/sq ft at 35 g/ℓ to 5 lb/day/sq ft at 70 g/ℓ. The high TSL value for test #708 is obviously out of line.

Comparison of Pilot Plant and Laboratory Results

For the first two plant operating periods in Table 7-7, an average TSL of 6.5 lb/day/sq ft was experienced. An average predicted underflow solids concentration of 33 g/ℓ was found for all laboratory tests shown in Table 7-8. This value compares well with that shown for the second pilot plant time period in Table 7-7. It should be recognized, however, that the average feed solids for the laboratory test data was 7.7 g/ℓ compared to over 11 g/ℓ for the pilot plant.

When polyelectrolyte was being used, laboratory test results predicted that maximum underflow would be achieved at the pilot plant TSL. Pilot plant results in Table 7-7 indicate maximum predicted underflow solids concentration was achieved. For a one on one comparison of 18 days of results, the range of pilot plant to laboratory test underflow solids concentration was from 0.2 to 1.0. In general, because of limited range of data compiled, no useful correlation was found between pilot plant and laboratory test results.

Summary Discussion

Pilot plant operating experience indicated that, at TSL's of 4 to 9 lb/day/sq ft and sludge SRT's of 1 to 4 days, no septicity problems were encountered.

Both pilot plant and laboratory test results indicate that the addition of polyelectrolyte to the chemical unit had a dramatic effect on underflow solids concentration. Since no meaningful correlation was found between pilot plant and laboratory results, it is impossible to suggest a definitive scale-up factor for the laboratory test results. It would seem reasonable to expect a prototype thickener to produce near maximum underflow solids concentration of 50+ g/ℓ from a feed solids concentration of about 24 g/ℓ at a TSL of 10 to 15 lb/day/sq ft.

Spent Carbon Sludge Thickening

All spent carbon produced during this study was thickened in a 5-ft diameter by 5-ft deep pilot plant gravity thickener. The sludge rakes were rotated at 16 rph resulting in a rake tip speed of about 4 fpm.

There were three general types of spent carbon sludges produced. The first was once used virgin carbon with lime pretreatment. The second was once used virgin carbon with alum or ferric chloride pretreatment. The third type was spent regenerated carbon with alum or ferric chloride pretreatment.

The once used virgin carbon had 7 to 15 percent volatile solids, 7 to 17 percent ash and 69 to 86 percent fixed carbon. The spent regenerated carbon had about 13 percent volatile solids, 21 percent ash, and 66 percent fixed carbon. The volatile solids were presumably composed of adsorbed organics and biologically generated solids.

The ash material was a combination of original ash (about 3 percent), chemical solids from pretreatment and for regenerated carbon some fine sand (less than 60 U.S. Mesh) from the fluidized bed furnace.

Data from some 32 laboratory gravity settling tests indicated that carbon sludge (slurry) density was 1.000 plus 0.00017 times g/ℓ of suspended solids. The term 0.00017 is about one-half of what would be expected for pure activated carbon solids with a particle density of 1.5 g/cc. This lower value was obviously due to the presence of considerable alum or ferric floc and biological solids.

Pilot Plant Results

As will be shown later in this Section, the pilot plant thickener was grossly underloaded. Consequently the unit required little operator attention. Due to often times sporadic operation of the carbon regeneration system the thickener was frequently used to store thickened carbon.

Five reasonably stable thickener operation and performance periods were identified. The results for these periods are shown in Table 7-10. Scrutiny of the carbon contactor operating data indicated only one identifiable effect on blowdown solids concentration. Going from period 2 to 3 a drastic reduction in blowdown solids concentration of from 44 to 11 g/ℓ was experienced. The cause had to be the substantial decrease in carbon dosage or the increase in dirunal flow variation. Previous operating experience at moderate to low carbon dosages (150 to 75 mg/ℓ) and constant plant flow did not result in low blowdown solids concentrations.[1] It is concluded that the low blowdown carbon solids experienced during period 3 in Table 7-10 was due to the relatively high diurnal flow variation. There is no obvious reason for the lower than expected blowdown carbon solids shown for period 4.

Gravity thickener performance results shown in Table 7-10 indicate that consistently high blowdown carbon solids concentration was achieved for various operating conditions. Variations shown for feed carbon solids, TSL or sludge age (SRT) did not affect blowdown carbon solids concentration.

TABLE 7-10: SPENT CARBON SLUDGE THICKENING: AVERAGE PILOT PLANT RESULTS

PERIOD #					
Calendar Dates	11/01/72 to 11/30/72	4/14/73 to 4/30/73	5/5/73 to 6/7/73	6/10/73 to 8/03/73	10/22/73 to 11/08/73
Carbon Contactor:					
Flow:					
Average gpm/sq ft	0.76	0.65	0.55	0.52	0.60
Peak gpm/sq ft	0.90	0.75	1.04	0.82	~0.90
Peak:Minimum	2:1	2:1	3:1	2:1	2:1
Pretreatment Chem.	Lime	Alum	Alum	$FeCl_3$	Alum
Carbon Dosage, mg/ℓ	120	294	86	110	70
Turbine Speed, rpm[a]	25	33	38	30	25
Reaction Zone-SS, g/ℓ	7.8	8.6	6.6	8.4	11
Blowdown-SS, g/ℓ	47	44	11	18	58
Gravity Thickener:					
Solids Loading, lb/day/sq ft	3.8	7.8	1.9	2.4	1.7
Sludge Depth, ft	2-3	2-3	2-3	2-3	1-2
Solids Capture, %	99+	99+	98+	97+	99+
Blowdown-SS, g/ℓ	155	151	154	100-154	142
Approximate SRT, Days	5-8	2-4	10-15	5-10	5-10

[a] 27 inch Diameter Turbine

The very good solids capture achieved was due to the very low hydraulic loadings of less than 20 gpd/sq ft. Thickener overflow suspended solids ranged from 50 to about 300 mg/ℓ.

The high carbon SRT's shown resulted in some anaerobic activity within the thickener. No significant odor problems were experienced relative to thickener operation. When thickened stored spent carbon was mixed in the 500 gal inventory tank prior to sampling, gross "rotten egg" odors were experienced. Full scale plant operation would not require such a mixing operation.

Thickener solids versus depth profiles are shown in Figures 7-11, 7-12, and 7-13. It is seen that from 1 to 2 ft of depth was required to reach near maximum carbon solids concentrations. The profiles in Figure 7-13, for period 5 in Table 7-10, indicate that maximum possible concentrations may not have been achieved.

Laboratory Test Results

Table 7-11 presents results from 31 laboratory thickening tests. These tests were all conducted on grab-composite samples obtained when alum or $FeCl_3$ coagulants were used in pretreatment. Three tests were made during the period of the study when regenerated carbon was reused.

Figure 7-14 shows a general trend of higher maximum solids concentrations for increasing feed solids concentrations. The data are, however, quite erratic at low carbon feed solids values.

Figures 7-15, 7-16, and 7-17 show the interrelationship between predicted TSL and feed carbon solids for various underflow carbon solids values. It is obvious that the responses shown are different than for the alum-primary or lime-primary sludges previously discussed. The data show that as feed carbon solids concentrations decrease below 30 to 40 g/ℓ, that higher predicted TSL are indicated.

Evaluation of the data shown on Figures 7-15, 7-16, and 7-17 indicated no significant effect due to alum versus ferric pretreatment. However, it is obvious that regenerated carbon can be thickened at significantly higher TSL's to produce a given underflow carbon concentration.

Comparison of Pilot Plant and Laboratory Results

Because the pilot plant thickener was loaded at a low level during nearly all operating periods, a direct comparison of pilot plant performance with laboratory results is impossible.

FIGURE 7-11: CARBON SLUDGE THICKENING - THICKENER SOLIDS PROFILE

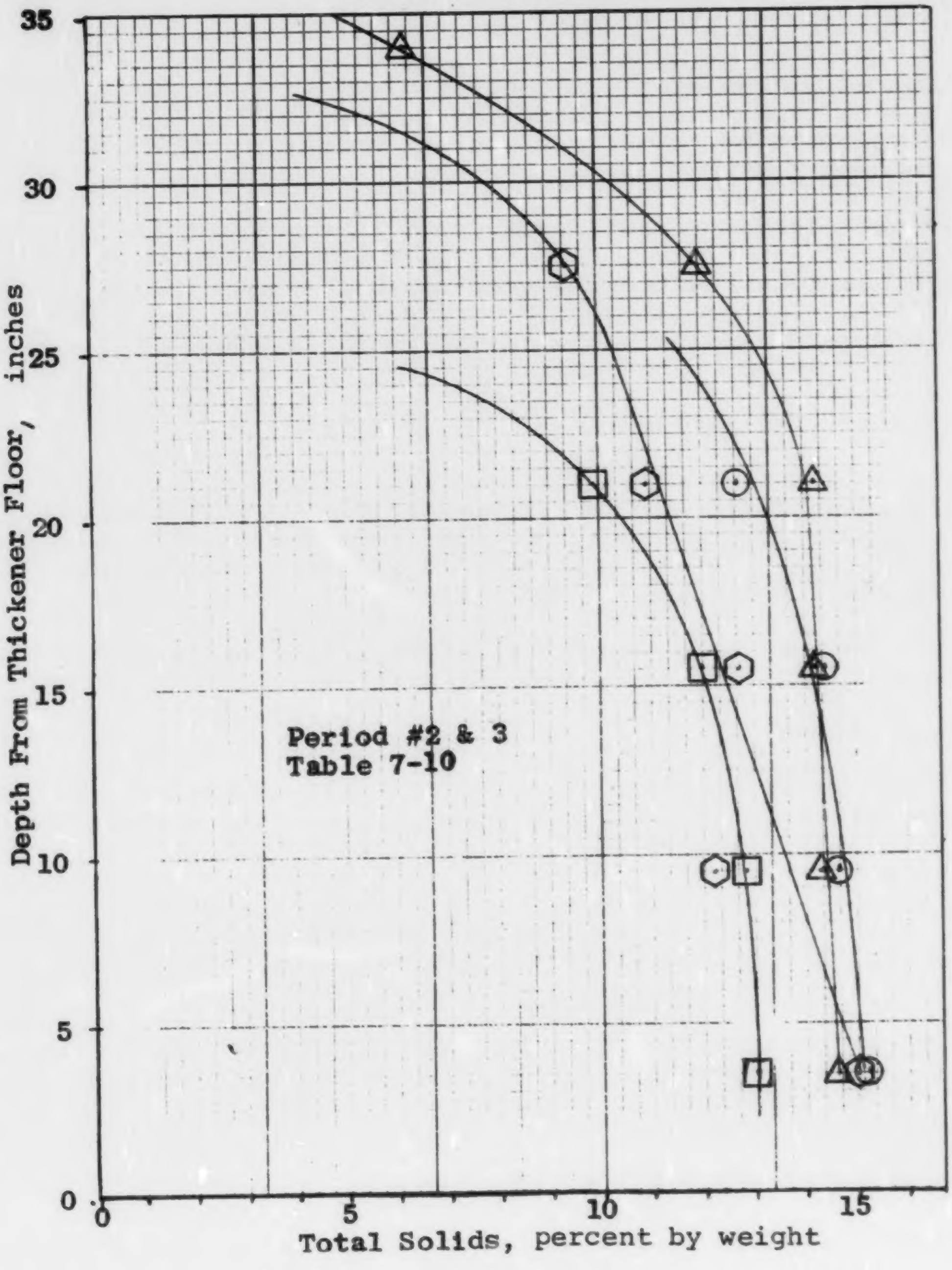


FIGURE 7-12: CARBON SLUDGE THICKENING - THICKENER SOLIDS PROFILE

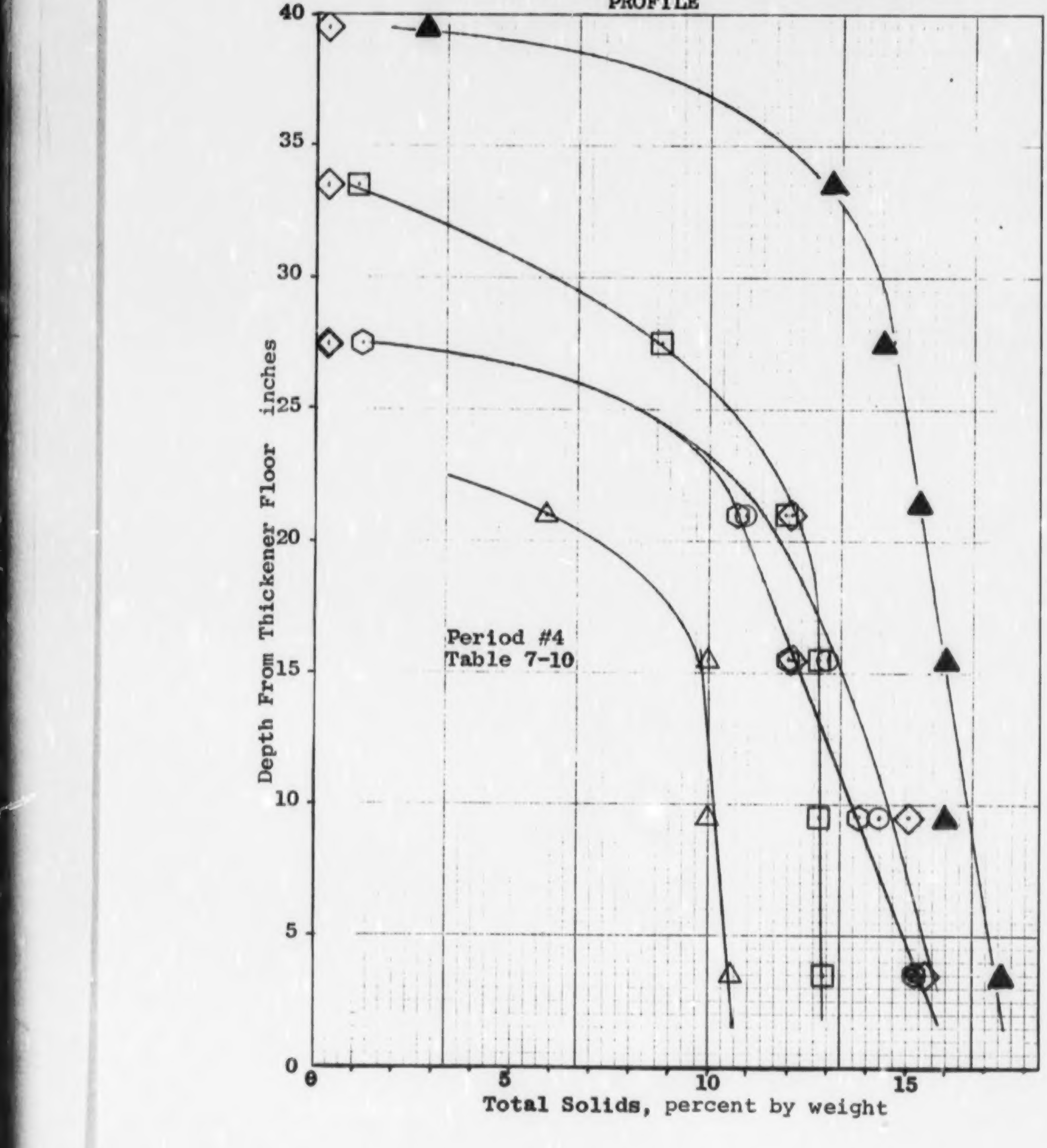


FIGURE 7-13: CARBON SLUDGE THICKENING - THICKENER SOLIDS PROFILE

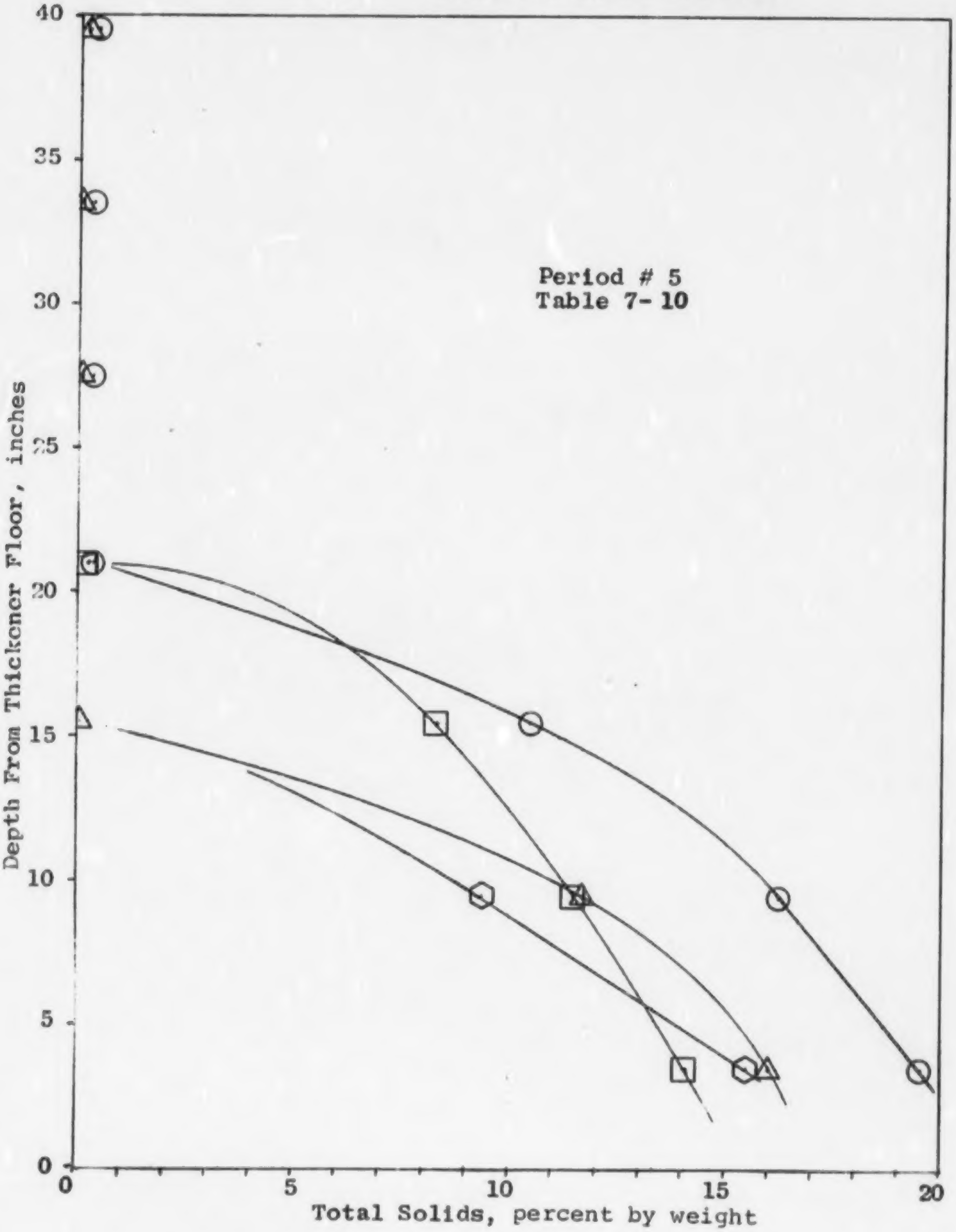


TABLE 7-11: CARBON LABORATORY GRAVITY THICKENING TEST RESULTS

Test #	Feed SS, g/ℓ	Thickener Solids Loading lb/day/sq ft @ Underflow SS (g/ℓ) of:							Max. Underflow SS, g/ℓ	Critical Point in inch/7 inch
		40	60	80	100	120	140	Max.		
621	6.0	64	51	47	44	--	--	43	120	2.1
803	7.3	82	52	44	--	--	--	40	104	1.4
512	8.2	46	34	30	--	--	--	30	82	2.0
610	8.5	20	15	--	--	--	--	13	76	1.5
529	10.6	94	54	50	--	--	--	41	96	1.7
619	13.2	60	30	24	--	--	--	20	106	1.5
510	14.2	--	67	55	48	45	--	44	135	2.0
520	14.3	--	--	54	50	--	--	50	95	2.1
701	18.2	107	51	32+	27	25	--	24	130	1.7
731	19.5	--	86	50	39	--	--	35	122	1.4
708	20.2	66	27	18	16	--	--	16	101	1.5
603	20.2	96	44	30	26	--	--	24	119	2.1
1025	21.8	53	21	15+	--	--	--	14	88	2.0
703	23.1	115	48	33	28	--	--	25	122	2.2
522	23.3	86	37	26	--	--	--	22	103	3.0
531	23.9	--	53	29	22	19	--	17	149	1.5
628	29.2	--	64	31	23	--	--	19	127	1.7
1106	28.8	--	63	39	31	27	--	25	135	2.9
426	30.2	--	35	17	12	--	--	11	113	1.8
424	35.3	--	40	19	13	--	--	10	123	2.4
1010	36.8	--	82	40	26	21	--	20	133	2.3
422	39	--	54	24	15	--	--	12	124	2.2
410	43.4	--	60	25	--	9.1	--	8.0	133	1.7
429	44.1	--	52	19	10.4	8.1	--	7.5	126	1.4
415	48.4	--	--	38	20	15	--	13	139	2.2

TABLE 7-11 (CONT.): CARBON LABORATORY GRAVITY THICKENING TEST RESULTS

Test #	Feed SS g/l	Thickener Solids Loading, lb/day/sq ft @ Underflow SS (g/l) of:					Max. Under-flow SS,g/l	Criti-cal Point Inch/7 Inch
		80	100	120	140	Max.		
419	51.8	40	20	15	--	13	130	2.4
1030	57.6	94	45	26	20	17	161	2.3
626	64.6	29	14	10.4	--	9.3	143	1.5
624	66.9	94	24	13	--	11.3	126	2.1
417	76.9	--	70	26	16	11	179	1.9
401	94.5	--	--	38	19	12	182	2.0

FIGURE 7-14: CARBON SLUDGE THICKENING - MAXIMUM UNDERFLOW CONCENTRATION

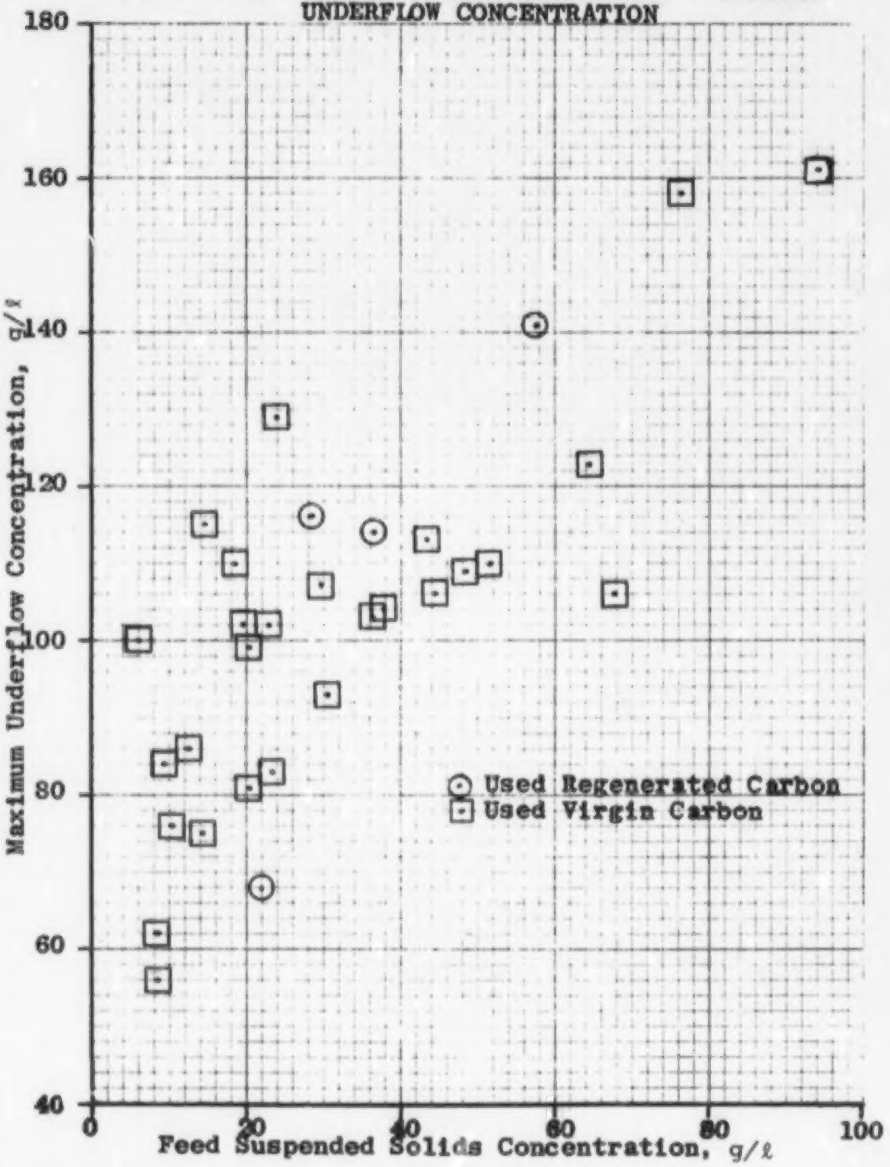


FIGURE 7-15: CARBON SLUDGE THICKENING - LABORATORY PREDICTED TSL VERSUS FEED SUSPENDED SOLIDS CONCENTRATION @ 120 g/ℓ UNDERFLOW CONCENTRATION

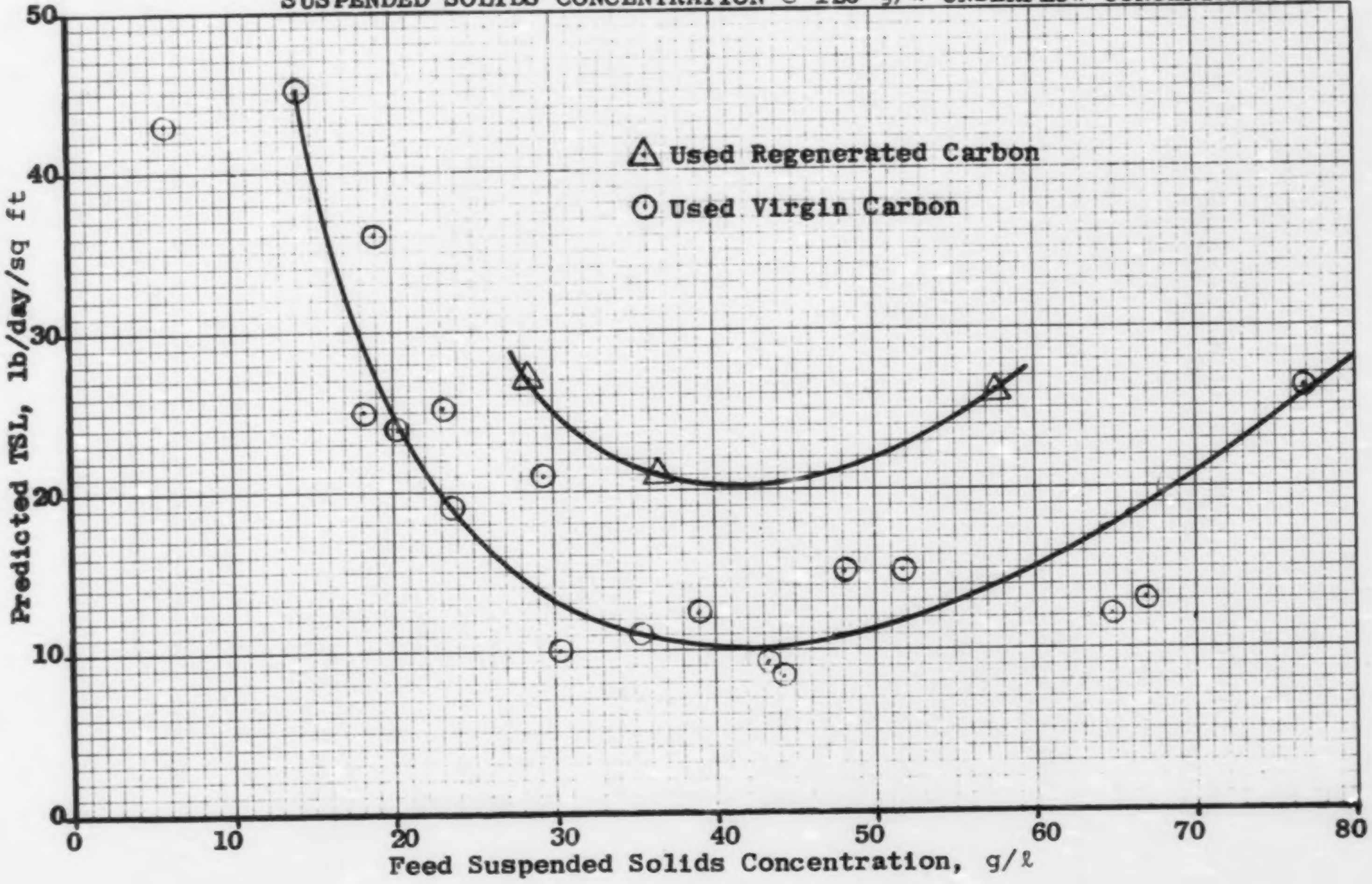


FIGURE 7-16: CARBON SLUDGE THICKENING - LABORATORY PREDICTED TSL VERSUS FEED SUSPENDED SOLIDS CONCENTRATION @ 140 g/ℓ UNDERFLOW CONCENTRATION

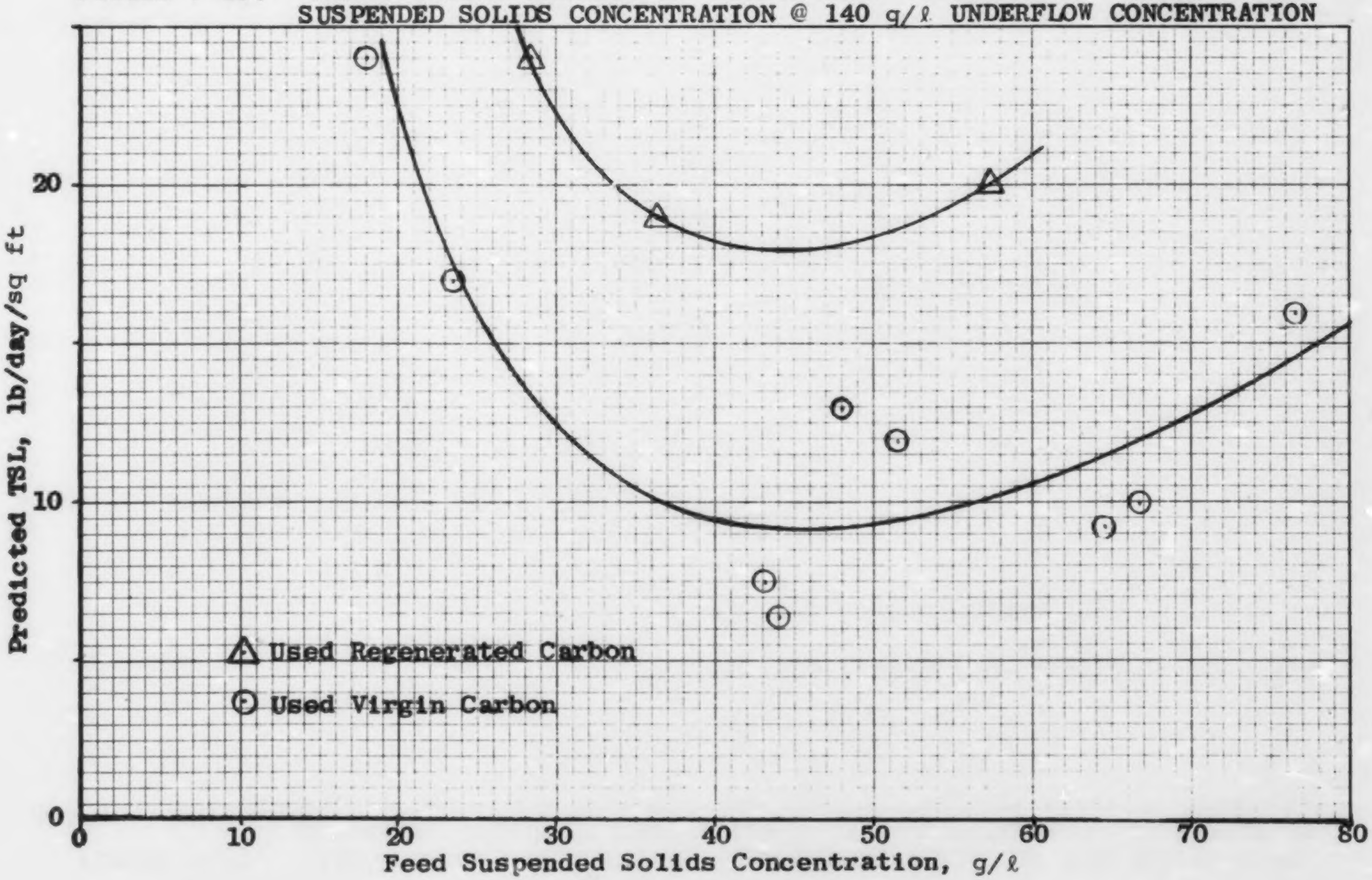
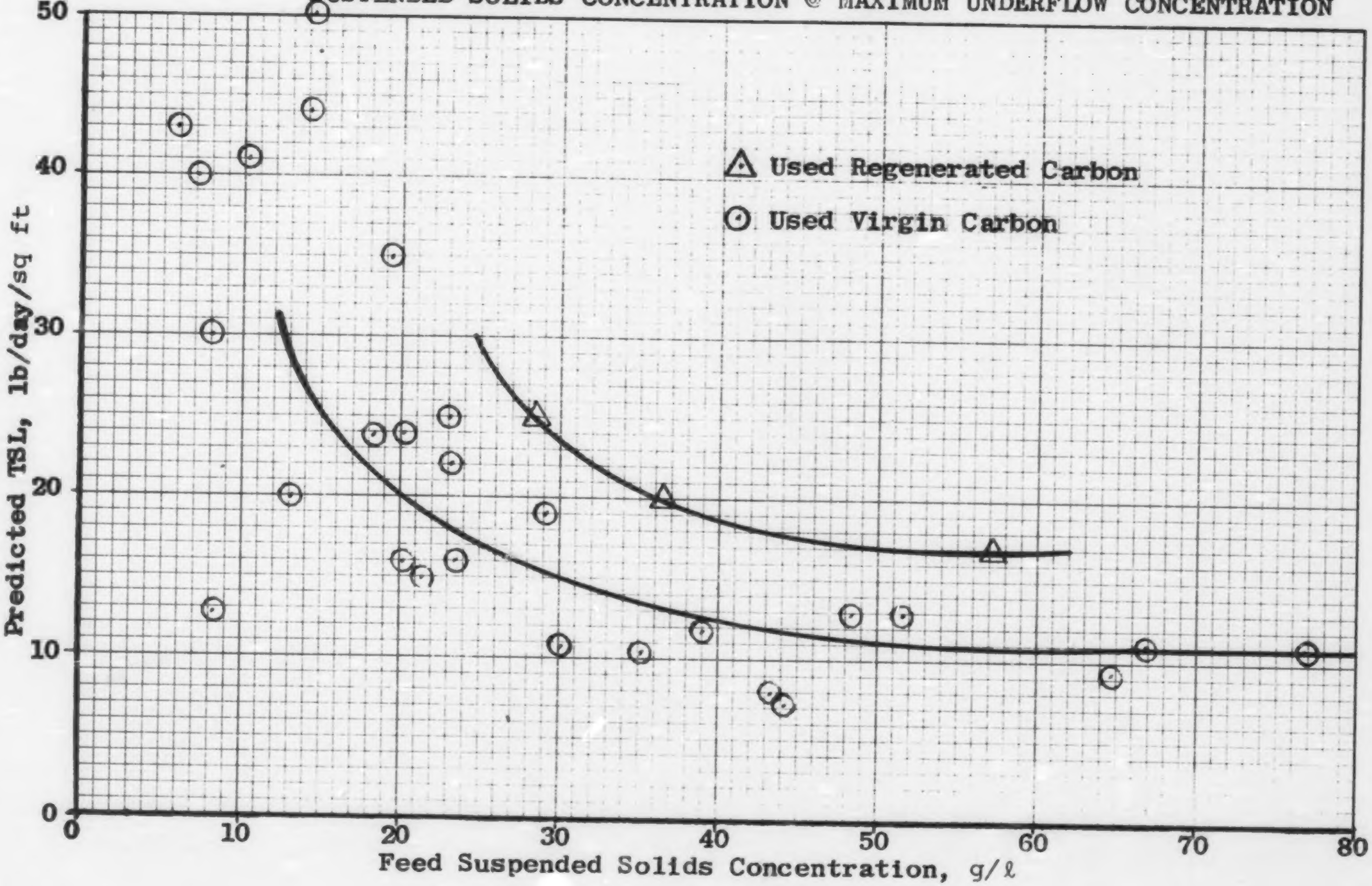


FIGURE 7-17: CARBON SLUDGE THICKENING - LABORATORY PREDICTED TSL VERSUS FEED SUSPENDED SOLIDS CONCENTRATION @ MAXIMUM UNDERFLOW CONCENTRATION



The highest pilot plant loading indicated in Table 7-10 was 7.8 lb/day/sq ft. At this TSL, the laboratory results in Figures 7-14 and 7-17 indicate that a maximum underflow solids concentration of about 130 g/ℓ should have resulted for the 44 g/ℓ feed carbon solids. The predicted 130 g/ℓ is significantly less than the 151 g/ℓ experienced.

In general, pilot plant underflow carbon solids concentrations experienced were well in excess of laboratory predicted values.

Summary Discussion

Laboratory test results indicate that at a TSL of 10 lb/day/sq ft, a maximum underflow carbon solids concentration should be produced regardless of feed carbon concentration. Pilot plant results indicate that this underflow carbon concentration could be as high as 150 g/ℓ. A prototype gravity thickener designed at a TSL of 10 lb/day/sq ft with a 3-ft deep thickened sludge layer would result in a carbon SRT of less than 3 days. This relatively low SRT should not result in significant odor problems.

VACUUM FILTRATION

Gravity thickened sludges were intermittently dewatered on the 3-ft diameter by 3-ft face belt filter described in Section IV. The general approach to operation of the filter station, including routine sampling, is presented in Section V. The primary objectives of the vacuum filtration dewatering studies were to determine required chemical conditions, the maximum filter yield and minimum cake moisture obtainable for the sludges dewatered. A secondary objective was to determine the correlation between pilot plant and laboratory leaf test results.

Lime-Primary Sludge Filtration

Laboratory leaf test experience during the previous contract eliminated the need for determining the type of media to be used for this sludge. Polypropolene - 852F (POPR-852F)* was used for all tests. A 0.10 sq ft laboratory circular leaf was outfitted with new media at the same time a new media was installed on the pilot plant. This leaf was used for all pilot plant simulation studies.

During the first three vacuum filter runs erratic vacuum was experienced. The vacuum filter bridge block was removed and inspected. It was noted that the bridge block openings were improperly located resulting in vacuum being applied before a sector was completely submerged. A properly machined bridge block was installed; this eliminated the problem of fluctuating vacuum.

It should be noted that the drum submergence was 33 percent, when the sludge level was at 1/4 inch above the lower extremity of overflow pipe. The cake form time (FT) was less than 33 percent of the cycle time and was determined by the location of bridge block openings and drum submergence.

Six pilot plant vacuum filter runs were conducted. Twelve series of leaf tests were conducted. Six were for simulation of pilot plant operating conditions and six were to obtain background data on high feed solids concentrations and the use of polyelectrolyte conditioner.

Pilot Plant Results

Table 7-12 presents pertinent operating and performance data for lime-primary sludge vacuum filtration. The cake form

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*Media designations are those of Eimco BSP Division of Envirotech Corporation, Salt Lake City, Utah.

TABLE 7-12: LIME-PRIMARY SLUDGE: VACUUM FILTRATION PILOT PLANT RESULTS

Run #	1110-1	1110-2	1110-3	112-1	112-2	112-3	112-4	112-5	116-1
Operating Conditions:									
Form Time(FT), min[a]	1.46	2.01	2.67	2.86	2.86	2.16	1.46	1.15	3.60
Dry Time(θ_D), min	2.13	2.97	3.98	4.25	1.03	3.19	2.17	1.71	1.3
Vacuum, in. of Hg	17	16	15	7-15	12-20	7-15	7-15	9-17	10-18
Solids:									
Feed, g/ℓ	111	111	111	113	113	113	113	113	91
Filtrate, g/ℓ	0.38	0.34	0.40	0.58	0.61	0.48	0.48	0.41	1.3
Filter Cake:									
Thickness, in.	3/32	3/32	1/8	3/16	3/16	7/32	1/8	3/32	1/16
Discharge Code[b]	F-P	F-P	G	E	G	E	G	F-G	E
Dry Weight(W), lb/sq ft	0.216	0.232	0.295	0.348	0.322	0.332	0.234	0.201	0.260
θ_D/W, min sq ft/lb	9.9	12	14	12	3.2	9.6	9.3	8.5	5.0
Performance:									
Form Filtration Rate (FFR) lb/hr/sq ft	8.88	6.93	6.63	7.30	6.76	9.22	9.62	10.5	4.33
Cake Moisture, %	64	67	65	69	72	67	66	71	74
Solids Capture, %	99.6	99.6	99.6	99.5	99.5	99.5	99.5	99.5	98.6

[a] Form Time/Cycle Time = 0.287 through Run #116, then = 0.26

[b] See Table 7-13 for Code

TABLE 7-12 (CONT.): LIME-PRIMARY SLUDGE: VACUUM FILTRATION PILOT PLANT RESULTS

Run #	116-2	118-1	118-2	118-3	118-4	222-1	222-2	222-3	222-4
Operating Conditions:									
Form Time(FT), min	2.08	2.91	2.81	1.73	0.92	2.72	2.45	2.12	1.34
Dry Time(θ_D), min	3.1	4.40	0.90	2.63	1.39	4.08	3.70	3.20	2.02
Vacuum, in. of Hg	7-13	14-	17+	15	15-	11	11+	12-	13+
Solids:									
Feed, g/ℓ	91	116	116	116	116	62	71	71	72
Filtrate, g/ℓ	0.82	0.29	0.23	0.32	0.30	--	--	--	--
Filtrate Turbidity, JTU	--	--	--	--	--	120	42	42	21
Filter Cake:									
Thickness, in.	1/16	1/8	1/8	7/64	3/32	1/8-	3/16	3/16	1/8
Discharge Code	G	E	G	F-	G	G	G+	G+	G-
Dry Weight(W), lb/sq ft	0.196	0.268	0.281	0.249	0.204	0.181	0.272	0.271	0.223
θ_D/W, min lb/sq ft	16	16	3.2	11	6.8	23	14	12	9.1
Performance:									
Form Filtration Rate (FFR) lb/hr/sq ft	5.65	5.52	6.00	8.64	13.30	3.99	6.66	7.67	9.99
Cake Moisture, %	74	72	74	69	71	75	74	73	73
Solids Capture, %	99.1	99+	99+	99+	99+	99+	99+	99+	99+

TABLE 7-12: (CONT.): LIME-PRIMARY SLUDGE: VACUUM FILTRATION PILOT PLANT RESULTS

Run #	222-5	222-6	222-7	222-8	223-1	223-2	223-3	223-4
Operating Conditions:								
Form Time(FT), min	1.39	1.28	.78	.79	2.73	1.39	1.27	1.08
Dry Time(θ_D), min	2.1	1.94	1.18	1.19	4.13	2.10	1.93	1.63
Vacuum, in. of Hg	13	14-	13	14+	8+	11-	14	16
Chem. Dose,[a] lb/TDS	--	--	1.0	1.8	0	1.2	2.8	4.9
Solids:								
Feed, g/ℓ	72	72	72	60	44-	44	44	44
Filtrate Turbidity JTU	15	15	15	23	61	78	47	77
Filter Cake:								
Thickness, in.	5/32	1/8	3/32	1/8-	1/8	3/32+	1/8	1/8
Discharge Code	--	--	--	--	F	F	G+	G+
Dry Weight(W), lb/sq ft	0.290	0.246	0.201	0.271	0.160	0.159	0.186	0.197
θ_D/W, min. lb/sq ft	7.2	7.9	5.9	4.4	26	13	10	8.3
Performance:								
Form Filtration Rate (FFR) lb/hr/sq ft	12.5	11.5	15.5	20.6	3.52	6.86	8.79	10.9
Cake Moisture, %	73	73	74	75	78	78	76	76
Solids Capture, %	99+	99+	99+	99+	99+	99+	99+	99+

[a] Anionic Polyelectrolyte

times (FT's) shown were computed from the measured drum cycle time (CT), minute/revolution, and drum submergences. FT, dry time (θ_D), and CT relationships are shown as footnotes in Table 7-12. Filter dry cake weight (W) and moisture content (M_C) were determined on cake samples taken with a 0.0845 sq ft area circular "cookie cutter." Values reported are normally the average of two weight determinations with 2 to 4 "cookie cutter" samples per determination. Samples were taken across the drum face at the mid-sector point and/or along the drum circumference within a given sector.

The quality of filter cake discharge, shown in Table 7-12, was noted by operating personnel according to the code presented in Table 7-13.

The vacuum filter solids handling capacity in this subsection is reported as FFR and expressed as lb of dry cake solids (dried 105°C) per hour per sq ft. Full scale filter yield (Y) is related to FFR by the following expression:

$$Y = (FFR)\ (FT/CT)\ (SUF)$$

where FT/CT represents effective drum submergences and SUF is a suitable "scale-up factor" or possibly a "safety factor." By reporting results as FFR the sometime variation in effective submergence is factored out.

Inspection of Table 7-12 indicates that, in general, filter cakes of 1/8 inch thickness or more exhibited "good" discharge. There was a "doctor blade" mounted just below the discharge roller, but its use was not judged to be a necessity for effectively discharging cakes of 1/8 inch or more thickness.

Figure 7-18 shows form filtration rate (FFR) as a function of cake FT. Theoretically for a uniformed sized particle suspension the slope of these lines would be -0.50. The results for feed solids concentration of 111 to 116 g/ℓ were found to have a -0.59 slope. The results for a feed solids concentration of 72 g/ℓ and 1.0 to 1.4 lb/ton of dry sludge solids (TDS) of anionic polyelectrolyte are seen to have a -0.85 slope.

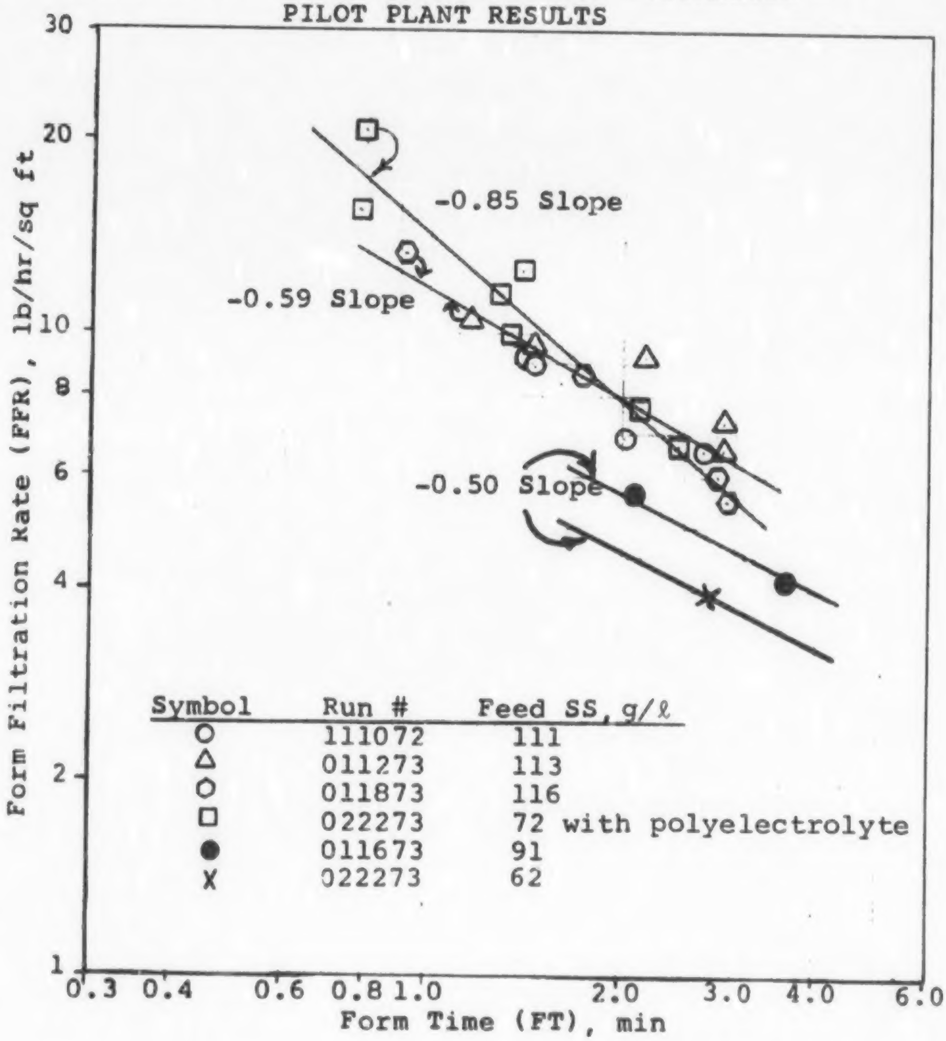
The data in Figure 7-18 show a increase in FFR for increasing feed solids concentration.

It is also apparent that a feed with 60-71 g/ℓ dewatered at rates similar to those for a feed with 111-116 g/ℓ, if 1.0 to 1.4 lb/TDS of anionic polyelectrolyte was used. The polyelectrolyte used was AtlaSep 2A2.

TABLE 7-13: ARBITRARY FILTER CAKE DISCHARGE CODE

Symbol	Designation	Number Weight
B+	Beautiful	16
B	Beautiful	15
B-	Beautiful	14
E+	Excellent	13
E	Excellent	12
E-	Excellent	11
G+	Good	10
G	Good	9
G-	Good	8
F+	Fair	7
F	Fair	6
F-	Fair	5
P+	Poor	4
P	Poor	3
P-	Poor	2
X	Wouldn't Discharge	1

FIGURE 7-18: LIME PRIMARY SLUDGE FILTRATION - PILOT PLANT RESULTS



The filter cake M_C was also found to be affected by feed solids concentration. Figure 7-19 shows M_C as a function of a "simplified correlating factor," θ_D/W. At θ_D/W greater than 10 min.sq ft/lb, minimum M_C's were achieved. The M_C data at this θ_D/W value also indicates that a decrease of about 2 percent M_C was experienced for an increase in feed solids of about 40 g/ℓ.

The 70-90 g/ℓ feed solids data shown in Figure 7-19 indicates that the use of polyelectrolyte did not result in any substantial increase in filter cake M_C.

Leaf Test Results

Each time the pilot plant vacuum filter was operated, a leaf test was conducted at near identical operating conditions of FT, θ_D, and vacuum level of a grab-composite of pilot plant feed. Table 7-14 presents the ratio of pilot plant to leaf test results. It is readily apparent that a high degree of correlation between the two existed. The average ratio of pilot plant to leaf test FFR is seen to be 1.15. For those runs without polyelectrolyte, the average and standard deviation was found to be 1.22 and 0.30, respectively. For the tests with polyelectrolyte, the average and standard deviation was 1.03 and 0.15. The more precisely correlated results with polyelectrolyte may have occurred because these runs were conducted on two consecutive days by the same personnel.

The average and standard deviations of the ratios of pilot plant to leaf test cake M_C for all data and those runs without polyelectrolyte were identical at 0.98 and 0.025. The results with polyelectrolyte were more precise with an average of 0.99 and a standard deviation of 0.016. The correlation of pilot plant and leaf test cake thickness was very good at 0.99.

The high degree of correlation between pilot plant and leaf test results indicated in Table 7-14 indicated that leaf test results could be used confidently to predict pilot plant performance.

As discussed in the lime-primary sludge thickening part of this Section, a more concentrated pilot plant underflow SS concentration was expected. Evaluation of previous contract work had indicated that attainment of thickened lime-primary sludge of 180 g/ℓ should be obtainable.[1] As seen in Table 7-12, the highest vacuum filter feed solids concentration experienced was 116 g/ℓ. To extend the range of results for

FIGURE 7-19: LIME PRIMARY SLUDGE FILTRATION, PILOT PLANT RESULTS - EFFECT OF FEED SUSPENDED SOLIDS CONCENTRATION ON CAKE PERCENT MOISTURE

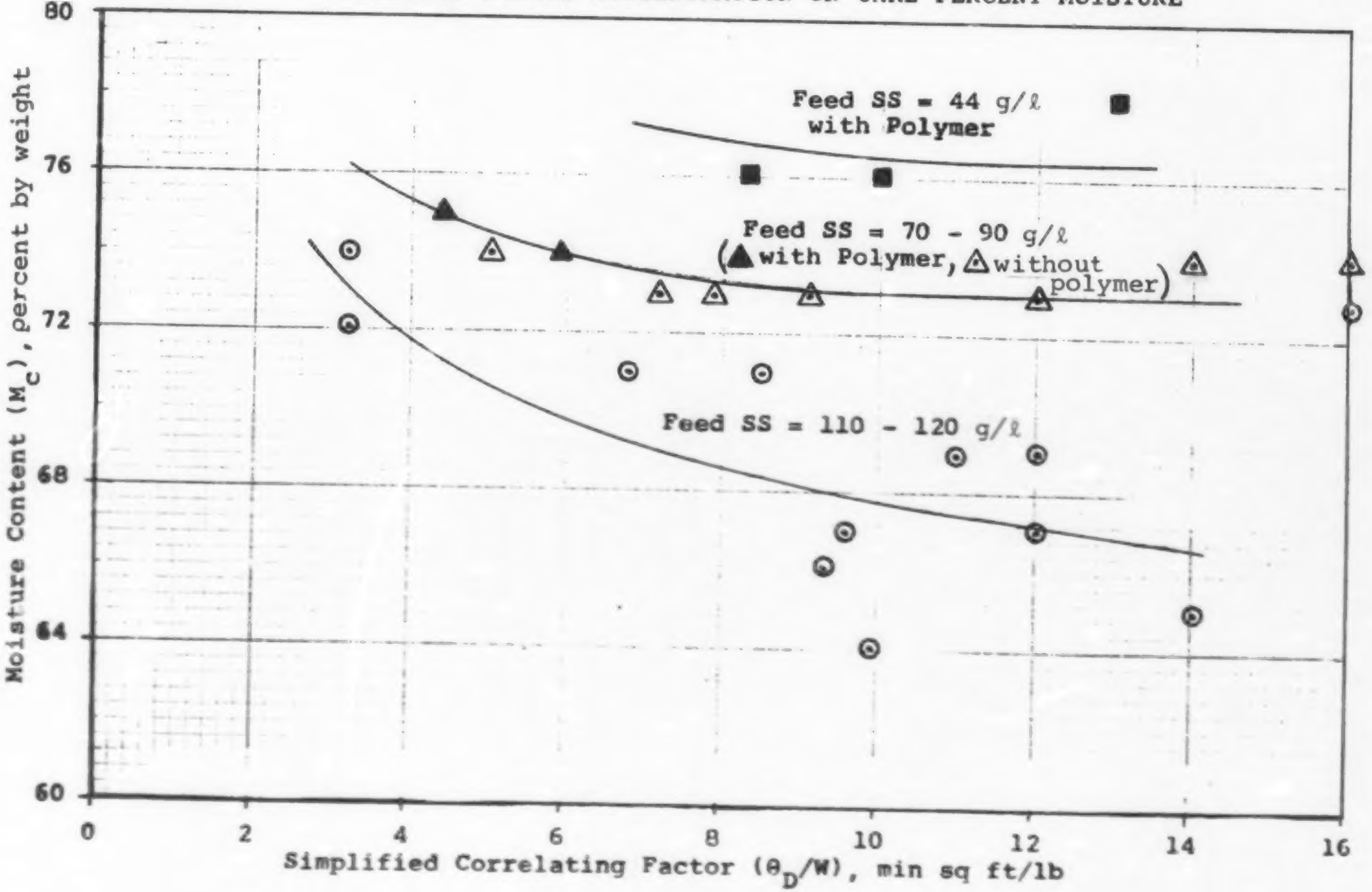


TABLE 7-14: LIME-PRIMARY SLUDGE FILTRATION: RATIO OF PILOT PLANT TO LEAF TEST RESULTS

Run #	Form Filtration Rate	Cake Moisture Content	Cake Thickness
1110-1	1.27	0.970	1.00
1110-2	1.40	0.971	1.00
1110-3	1.62	0.956	1.14
112-1	1.11	0.972	1.00
112-2	0.87	1.000	1.00
112-3	1.19	0.971	1.17
112-4	1.05	0.904	1.00
112-5	1.13	0.973	0.75
116-1	1.41	0.949	0.50
116-2	1.77	0.974	1.00
118-1	1.18	1.014	1.00
118-2	0.98	1.000	0.67
118-3	1.68	0.986	1.17
118-4	0.89	1.014	0.86
222-1	0.91	1.000	1.00
222-2P	0.88	0.987	1.00
222-3P	0.87	0.987	1.00
222-4P	1.00	1.000	1.33
222-5P	1.18	0.987	1.25
222-7P	0.94	1.014	1.00
222-8P	1.31	1.000	1.00
223-1	1.01	0.975	1.00
223-2P	1.14	0.987	1.00
223-3P	0.90	1.013	1.00
223-4P	1.08	0.962	1.00
Average	1.15	0.983	0.994
Std. Deviation	0.26	0.025	0.169

the current study, leaf tests were conducted at higher feed solids concentrations. In addition, the effect of polyelectrolyte on filter yields was evaluated by leaf tests.

Prior to vacuum filter run #222 (see Table 7-12) two series of leaf tests were conducted using polyelectrolyte conditioner. The result of these two tests are presented in Table 7-15. From the results for test series #214 it is seen that for a lime-primary sludge at about maximum concentration, polyelectrolyte has a substantial affect on the FFR. Extrapolation of data for test #214-3, indicates that, at a FT of 0.4 min, a FFR of about 100 lb/hr/sq ft could be achieved. Cake thickness would be in excess of 1/8 inch. For a conventional belt filter this would mean a yield in excess of 25 lb/hr/sq ft.

The results for test series #221 presented in Table 7-15 show a less dramatic and limited effect of polyelectrolyte on FFR. For test series #221, a feed solids concentration of 118 g/ℓ was diluted to 88 g/ℓ to accommodate the addition of various polyelectrolyte dosages. No consistent increase in FFR was experienced for polyelectrolyte dosages above 1.1 lb/TDS. In addition, the FFR at 1.1 lb/TDS was less than for the 118 g/ℓ feed solids concentration without polyelectrolyte addition. This result is similar to the results for pilot plant runs #222 and #223 in Table 7-12. It appears that the higher the feed solids concentration, the larger the increase in filter yield for a given polyelectrolyte dosage.

It is germane to note that the leaf test results for test #221, in Table 7-15, indicated no significant reduction in filter cake M_C with polyelectrolyte dosages up to 3.2 lb/TDS. Test #214 shows an increase in M_C. The reduction in feed solids concentration and θ_D/W values could be the reason for this observation and not the use of polyelectrolyte.

Two series of leaf tests were conducted at feed solids concentrations of 232 and 189 g/ℓ. The results of FFR versus FT are presented in Figure 7-20. The ratio of θ_D/FT during these tests was 1.5. The average filter cake M_C for these tests was 66 percent (62 to 70 percent range). Also shown on Figure 7-20 is the curve for pilot plant results at feed solids concentration of 111 to 116 g/ℓ. The pronounced effect of high feed solids concentration on FFR is obvious.

Summary Discussion

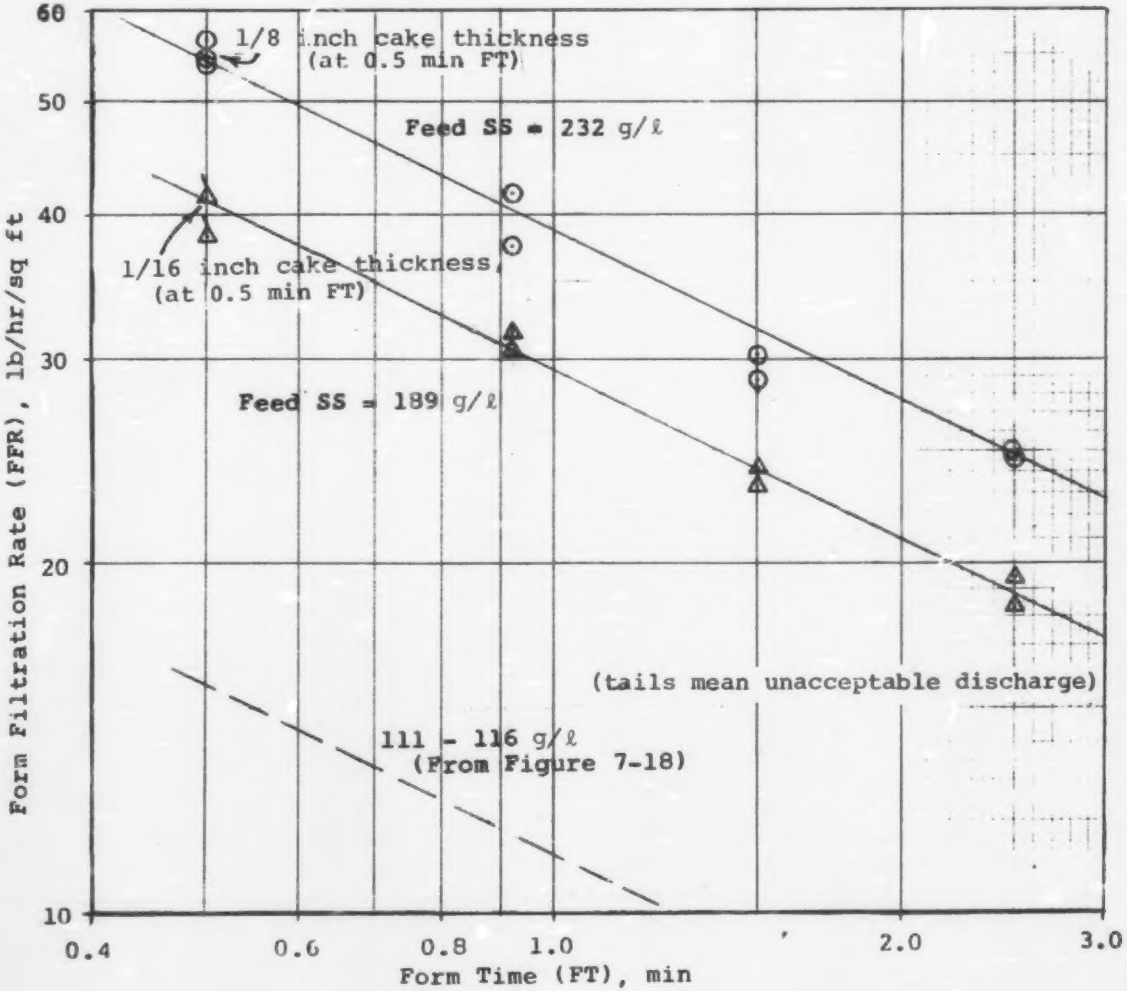
Lime-primary filter cakes of 1/8 inch thickness were found to exhibit "good" discharge characteristics. Filter

TABLE 7-15: LIME-PRIMARY SLUDGE: LEAF TEST RESULT USING POLYELECTROLYTE

Run #	214-1	214-2	214-3	221-1	221-2	221-3	221-4	221-5	221-6	221-7
Operating Conditions:										
Form Time(FT),min	1.50	1.50	1.50	1.50	1.50	1.50	1.50	1.50	1.50	1.50
Dry Time(θ_D),min	2.27	2.27	2.27	2.27	2.27	2.27	2.27	2.27	2.27	2.27
Vacuum, in.of Hg	20	20	20	20	20	20	20	20	20	20
Chem. Dose,[a] lb/TDS	0	0.56	1.24	0	0	1.1	2.3	2.7	3.2	4.5
Solids:										
Feed, g/ℓ	230	215	194	118	88	88	88	88	88	88
Filtrate, g/ℓ	--	--	--	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Filter Cake:										
Thickness, in.	1/4	5/16	7/16	1/8+	3/32	1/8	1/8+	3/32	1/16	1/4
Dry Weight (W), lb/sq ft	0.708	0.829	1.32	0.278	0.181	0.239	0.240	0.228	0.262	--
θ_D/W,min sq ft/lb	3.2	2.7	1.7	8.2	12.5	9.5	9.5	10.0	8.7	--
Performance:										
Form Filtration Rate (FFR) lb/hr/sq ft	28.3	33.2	52.8	11.1	7.24	9.56	9.60	9.12	10.5	--
Cake Moisture, %	61	63	66	71	73	72	73	73	74	--
Solids Capture, %	--	--	--	99+	99+	99+	99+	99+	99+	99+

[a] Anionic Polyelectrolyte

FIGURE 7-20: LIME PRIMARY SLUDGE FILTRATION, LEAF TEST RESULTS - EFFECT OF HIGH FEED SUSPENDED SOLIDS CONCENTRATION



cakes just under 1/8 inch were marginally discharged with the use of a "doctor blade".

The average of all pilot plant and leaf test data for W and thickness indicated a filter cake density of 23 lb of dry solids per cu ft of cake.

The effect of feed solids concentration on FFR over a broad range had to be determined by leaf tests. Figure 7-21 presents all leaf test results. The solid curve shown indicates the FFR expected for a filter cake thickness of about 1/8 inch and vacuums in the 11 to 17 inches of Hg range. This curve represents the maximum yield obtainable. The dashed curve is from the previous contract report.[1] Though cake thickness and vacuum are slightly different, the previous and current results are obviously similar. Of significance is the dramatic effect of feed solids concentration above 120 g/ℓ. The obvious implication of these results is that maximization of thickening prior to the dewatering step is very beneficial. The economic trade-offs between clarifier, thickener, and vacuum filter operation will be left for others to determine.

The significant effect of feed solids concentration on filter cake M_c is seen in a composite of leaf test data shown on Figure 7-22. At θ_D/W values above about 4 min.sq ft/lb, about 10 percentage points less M_c was experienced for feed solids concentrations of 189 to 232 g/ℓ compared to 111 to 116 g/ℓ. The economic impact of a 10 percentage point reduction of M_c on a subsequent incineration operation would be very substantial.

Alum-Primary Sludge Filtration

During the period of April to June, 1973, alum was used in the chemical pretreatment step. During this period, nine pilot plant vacuum filter runs were conducted on gravity thickened sludge. Leaf tests were conducted to simulate pilot plant performance and to determine the effects of several operating variables.

Preliminary Leaf Tests

Preliminary leaf tests were conducted to determine a suitable filter media and chemical conditioning to achieve a high degree of solids capture and good filter cake discharge. Figure 7-23 presents results from some of these tests. FFR's are reported on the basis of lbs of dry alum-primary solids per hour per sq ft. Scrutiny of these results indicated that NY-317F provided the highest FFR's

FIGURE 7-21: LIME PRIMARY SLUDGE FILTRATION, LEAF TEST RESULTS - EFFECT OF FEED SUSPENDED SOLIDS CONCENTRATION

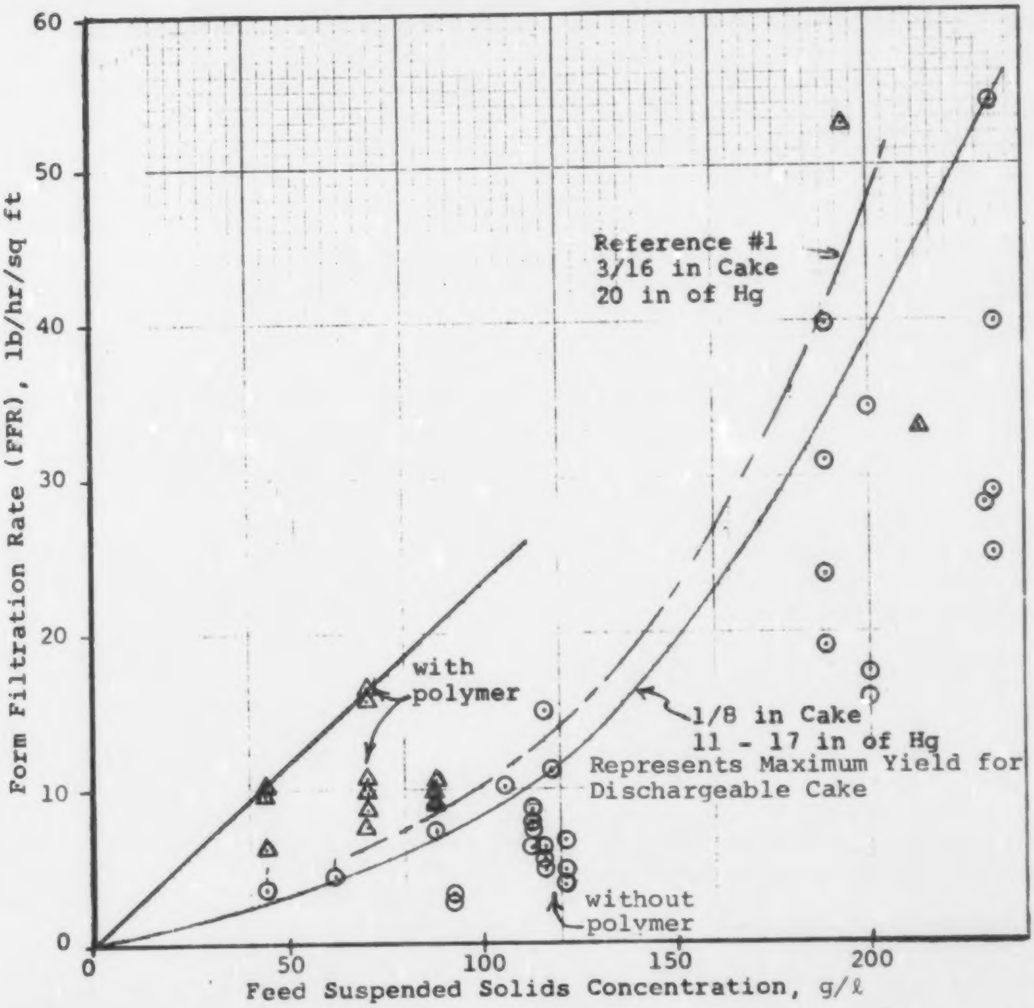


FIGURE 7-22: LIME PRIMARY SLUDGE FILTRATION, LEAF TEST RESULTS - CAKE MOISTURE CONTENT

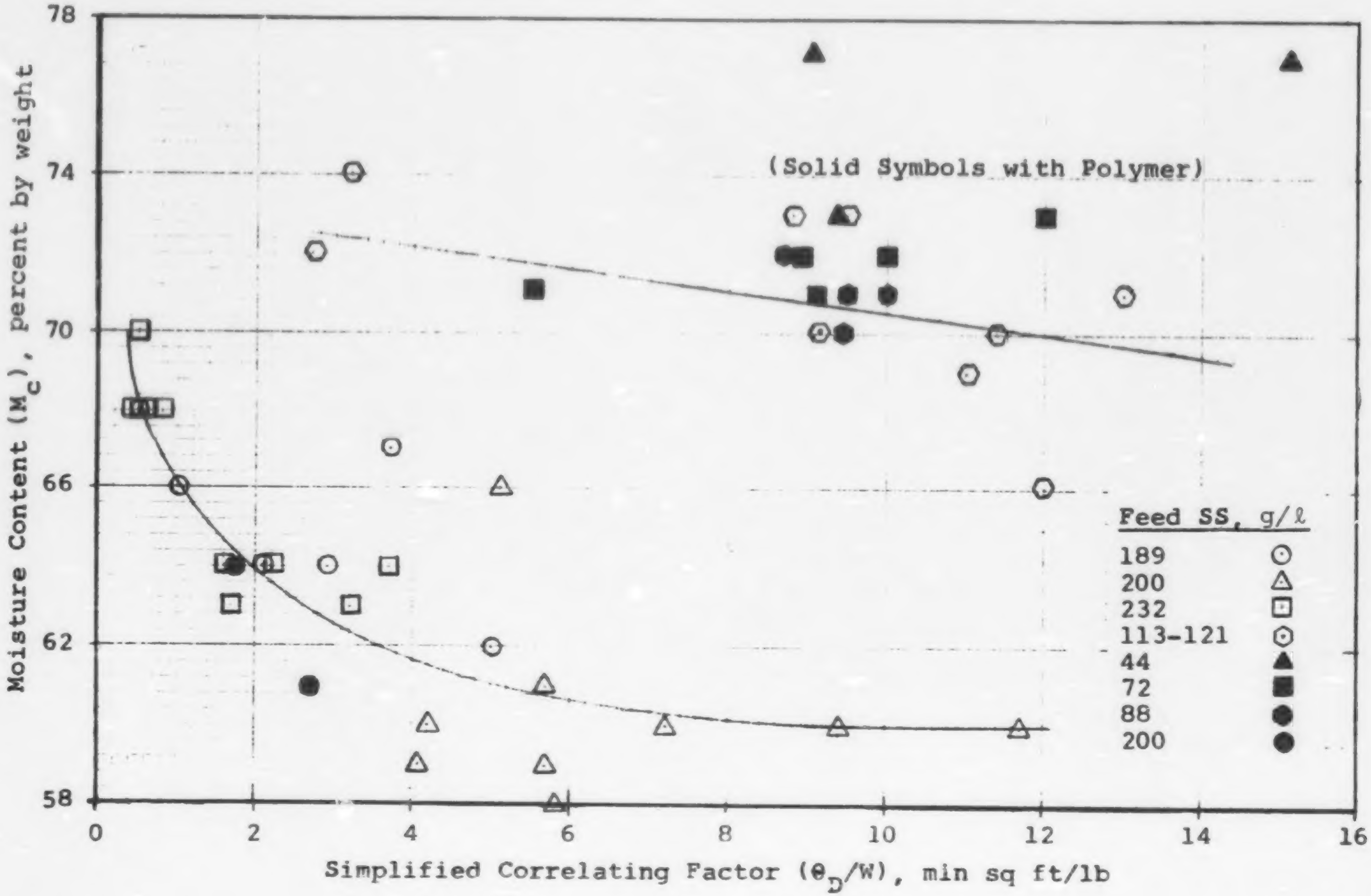
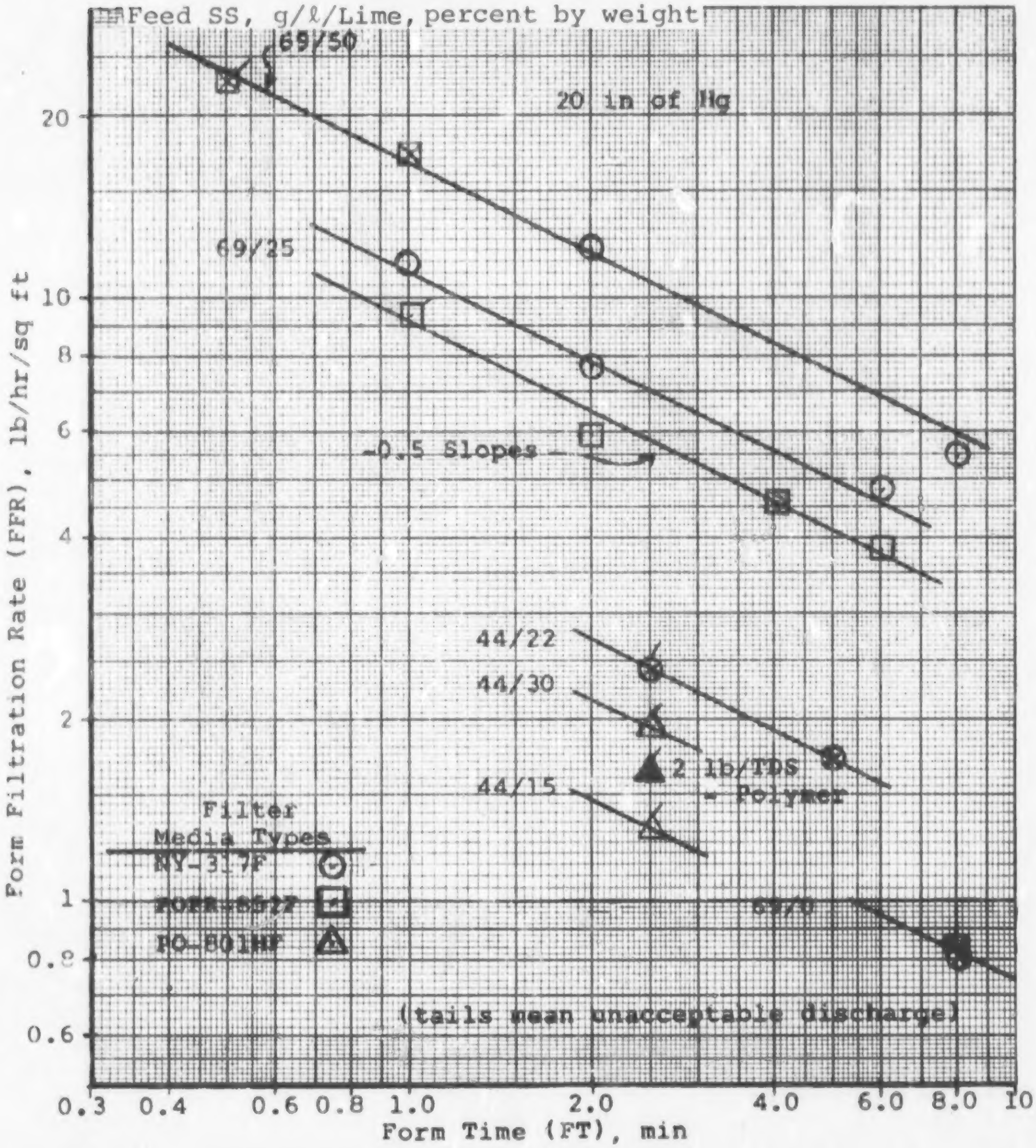


FIGURE 7-23: ALUM PRIMARY SLUDGE FILTRATION, LEAF TEST RESULTS - EFFECT OF MEDIA AND CHEMICALS



with the exception of the results for 50 percent lime conditioning. The NY-317F produced both excellent solids capture and very good discharge of thin (1/8 inch) filter cakes as opposed to the other two media. Thus, a NY-317F belt was obtained for the pilot plant vacuum filter.

The preliminary leaf tests indicated that at least 20 percent lime conditioning chemical was required to produce a dischargeable cake. Conditioning of a 44 g/ℓ feed with 2 lb/TDS anionic polyelectrolyte did not result in a dischargeable cake. It was decided to use lime as the conditioning chemical at dosages in excess of 20 percent by weight.

Pilot Plant Results

Table 7-16 presents results of alum-primary sludge vacuum filtration tests. Runs #419 and #426 were conducted on fairly septic alum-primary sludge. Sludges filtered during Runs #531 and #604 were produced during a period when about 0.2 mg/ℓ of anionic polyelectrolyte was used in the chemical contactor-clarifier unit.

Two problems were encountered during alum-primary sludge filtration studies. The first resulted from the fairly fragile nature of the vacuum filter nylon belt. The 3-ft diameter by 3-ft face vacuum filter thermoplastic drum was sent back to the supplier, just prior to this series of tests, to be dismantled and rebonded. Elimination of vacuum leaks which resulted in achieving less than full vacuum capacities during the lime-primary sludge filtration tests was the purpose. The vacuum filter drum was returned in a shabby condition with numerous burrs and sharp edges. Operating personnel removed most of the burrs and sharp edges, but a few remaining ones caused tears in the media. Data in Table 7-16 indicate that starting with Run #521, reduced solids captures were experienced. By Run #607 most holes had been patched and solids captures are seen to return to about 99 percent.

The second operational problem resulted from inadequate vat mixing conditions. Alum-primary sludges of 50 to 70 g/ℓ were much more viscous than were 200 g/ℓ lime-primary sludges. Lime conditioning of the alum-primary sludge produced large floc particles which rapidly settled in the filter vat. On one occasion a 50 percent difference in solids concentration was found for samples taken at a high and low vat level. It was often felt that the flocculator mixer had to be run at excessive speeds to maintain a uniform suspension in the flocculator overflow, where feed samples were obtained. In an attempt to achieve a more

TABLE 7-16: ALUM-PRIMARY SLUDGE: PILOT PLANT VACUUM FILTRATION RESULTS

Run #	419-1	419-2	419-3	419-4	419-5	426-1	426-2	426-3	427-1	427-2
Operating Conditions:										
Form Time(FT)[a],min	2.7	1.43	0.88	2.70	0.30	0.59	0.88	2.67	2.70	1.38
Dry Time(θ_D),min	4.09	2.16	1.34	4.09	0.46	0.88	1.32	4.03	4.09	2.08
Vacuum, in.of Hg	18+	18+	18+	18+	17+	20	20	20	20	20
Lime Dose, % by wgt	40	38	39	100	100	44	46	52	37	37
Sludge, pH	10	10	8	12	--	11	11	11	11	11
Solids:										
Feed, g/ℓ	73	73	73	73	73	60	60	60	53	53
Feed & Lime, g/ℓ	84	81	73	103	--	73	77	87	65	65
Filtrate, g/ℓ	0.32	0.18	0.29	0.12	--	0.37	0.16	0.16	--	--
Filter Cake:										
Thickness, in.	3/16-	1/8+	3/32	5/16	5/64	1/16	1/16-	1/8	5/32-	5/32-
Discharge Code,[b]	G	G	F	G	F	G	F	E	G	G
Dry Weight(W), lb/sq ft	0.246	0.162	0.121	0.384	0.102	0.121	0.0865	0.169	0.196	0.200
θ_D/W,min sq ft/lb	16.6	13.3	11.1	10.7	4.5	7.3	15	24	21	10
Performance:										
Form Filtration Rate (FFR)[c] lb/hr/sq ft	4.75	6.13	8.25	6.05	14	10.1	4.55	2.62	3.55	7.09
Cake Moisture, %	74	76	75	78	77	74	73	73	76	76
Solids Capture, %	99+	99+	99+	99+	99+	99+	99+	99+	--	--

[a] Form Time/Cycle Time = 0.26

[b] See Table 7-13

[c] FFR's are reported on the basis of Lbs of Dry Alum-Primary solids and do not include the lime conditioning solids.

TABLE 7-16 (CONT.): ALUM-PRIMARY SLUDGE: PILOT PLANT VACUUM FILTRATION RESULTS

Run #	427-3	427-4	516-1	516-2	516-3	521	525-1	525-2	531-1	531-2
Operating Conditions:										
Form Time(FT),[a]min	0.83	0.36	1.38	2.70	1.11	1.4	2.76	0.96	1.48	1.19
Dry Time(θ_D),min	1.26	0.54	2.09	4.09	1.67	2.1	4.17	1.45	1.99	1.59
Vacuum, in.of Hg	20	20	17	17	17	19	18	20	16	15
Lime Dose,% by wgt	37	37	36	36	10	47	55	38	20	20
Sludge, pH	11	10.8	9.7	10	9	--	11.8	11.7	9.9	10.1
Solids:										
Feed, g/ℓ	53	53	34	34	34	21	44	44	62	62
Feed & Lime, g/ℓ	65	65	43	43	38	28	62	62	71	71
Filtrate, g/ℓ	--	--	0.30	0.49	0.27	6.1	0.17	0.09	1.0	5.4
Filter Cake:										
Thickness, in	1/8-	1/16	3/64	3/64+	1/32-	1/32	1/4	5/32	3/32+	3/32
Discharge Code	G	G	G	G	G	?	E	E	--	--
Dry Weight(W), lb/sq ft	0.160	0.107	0.0817	0.118	0.0798	0.0856	0.344	0.186	0.142	0.110
θ_D/W,min sq ft/lb	5.2	5	26	35	21	25	12	7.8	14	15
Performance:										
Form Filtration Rate (FFR)lb/hr/sq ft	9.43	14.5	2.81	2.07	3.86	2.75	5.31	8.25	5.03	4.84
Cake Moisture, %	75	76	76	75	75	73	77	77	76+	76+
Solids Capture, %	--	--	99	99	99	72	99+	99+	98+	92

[a] Starting with Run #531, FT/CT = 0.28

TABLE 7-16 (CONT.): ALUM-PRIMARY SLUDGE: PILOT PLANT VACUUM FILTRATION RESULTS

Run #	531-3	531-4	531-5	604-1	604-2	604-3	607-1	607-2	607-3	607-4
Operating Conditions:										
Form Time(FT),min	2.02	2.01	0.94	2.7	2.7	2.7	3.02	1.54	0.93	2.98
Dry Time(θ_D),min	2.71	2.69	1.26	4.1	4.1	4.1	4.03	2.07	1.25	3.99
Vacuum, in.of Hg	19	19	19	21	21	21	21	21	21	19
Lime Dose,% by wgt	10-15	25-35	25-35	18	27	27	35	30	35	15
Sludge, pH	9.5	10.9	10.4	10.3	10.4	10.3	10.8	10.7	10.7	8.9
Solids:										
Feed, g/ℓ	62	62	62	59+	--	--	49	49	49	49
Feed & Lime, g/ℓ	66	72	66	63+	67	67	58	57	53	50
Filtrate, g/ℓ	1.0	2.3	7.5	12	4.3	7.2	0.25	0.12	0.19	0.17
Filter Cake:										
Thickness, in.	1/16	3/64	7/64	1/32	1/32-	1/32-	1/8	3/64	1/16	1/32
Discharge Code	--	--	--	--	--	--	G+	G	G-	P-
Dry Weight(W), lb/sq ft	0.151	0.188	0.156	0.0739	0.0856	0.106	0.186	0.118	0.0911	0.0782
θ_D/W,min sq ft/lb	19	15	8.1	55	48	39	22	18	14	51
Performance:										
Form Filtration Rate (FFR)lb/hr/sq ft	4.21	4.83	9.35	1.53	1.68	2.07	3.12	3.95	5.43	1.54
Cake Moisture, %	75	76+	76-	80	76	74	75	75	74	74
Solids Capture, %	98+	97	88	81	94	89	99+	99+	99+	99+

TABLE 7-16 (CONT.): ALUM-PRIMARY SLUDGE: PILOT PLANT VACUUM FILTRATION RESULTS

Run #	611-1	611-2	611-3
Operating Conditions:			
Form Time(FT), min	3.0	0.89	1.52
Dry Time(θ_D), min	4.00	1.19	2.03
Vacuum, in. of Hg	19	20	18-
Lime Dose, % by wgt	27	27	27
Sludge pH	10.7	10.8	10.7
Solids:			
Feed, g/ℓ	41	41	41
Feed & Lime, g/ℓ	48	48	48
Filtrate, g/ℓ	0.42	0.47	2.4
Filter Cake:			
Thickness, in.	1/8-	1/16	3/32
Discharge Code	G+	F-	G+
Dry Weight(W), lb/sq ft	0.152	0.086	0.119
θ_D/W, min sq ft/lb	26	14	17
Performance:			
Form Filtration Rate (FFR) lb/hr/sq ft	2.60	4.95	4.01
Cake Moisture, %	79	79	78
Solids Capture, %	99	99	99

uniform filter vat suspension, the vat overflow method was changed after Run #525. Sludge was fed to the vat at a rate in excess (about 25 percent) of what could be dewatered. The excess sludge overflowed the front edge of the vat into the belt wash trough and to drain. Thus sludge was fed at the center back of the vat, flowed underneath the drum and the excess flowed over the entire length of the front edge of the vat. This method of operation improved vat mixing somewhat.

Figure 7-24 shows FFR versus FT for several of the pilot plant runs presented in Table 7-16. The tails on some data points represent unacceptable cake discharge. The FFR's reported in Table 7-16 and subsequent graphs are on the basis of lbs of dry alum-primary sludge solids and do not include the lime solids.

Most of the data in Figure 7-24 are seen to fit the -0.5 slopes shown. The data is not extensive or consistent enough to define precise trends in the effect of feed solids or percent lime used. The data for the 60 g/ℓ feed solids of septic sludge is obviously quite low considering the approximate 50 percent lime was used. In general, higher feed solids and/or higher lime dosages resulted in higher FFR's.

The M_C's in Table 7-16 range from 73 to 80 percent. These values are reported as a percent of total filter cake solids, including lime solids. There is no obvious trend in the data of lower M_C for higher θ_D/W's. After presentation of leaf test results, M_C will again be discussed.

Leaf Test Results

Each pilot plant test was simulated with a leaf test. The ratio of pertinent pilot plant to leaf test performance parameters are presented in Table 7-17. Scanning the results and oberving the standard deviations indicates that the ratios are quite variable. After the previously noted change in filter vat overflow method, Runs #531 to #611, the variability of the data was reduced slightly.

A probable cause of some of the variation was found to be due to the time lag between sample procurement and conducting the leaf test. Simulation leaf tests for Run #419 were conducted after 1 day and for Run #516 after 6 days time lag. The erratic ratios of pilot plant to leaf test performance for these and other runs prompted an evaluation of of time lag on FFR. Figure 7-25 shows results for several leaf tests. The effect of time between sample procurement and conducting the leaf test is seen to be inconsistent,

FIGURE 7-24: ALUM PRIMARY SLUDGE FILTRATION - PILOT PLANT RESULTS

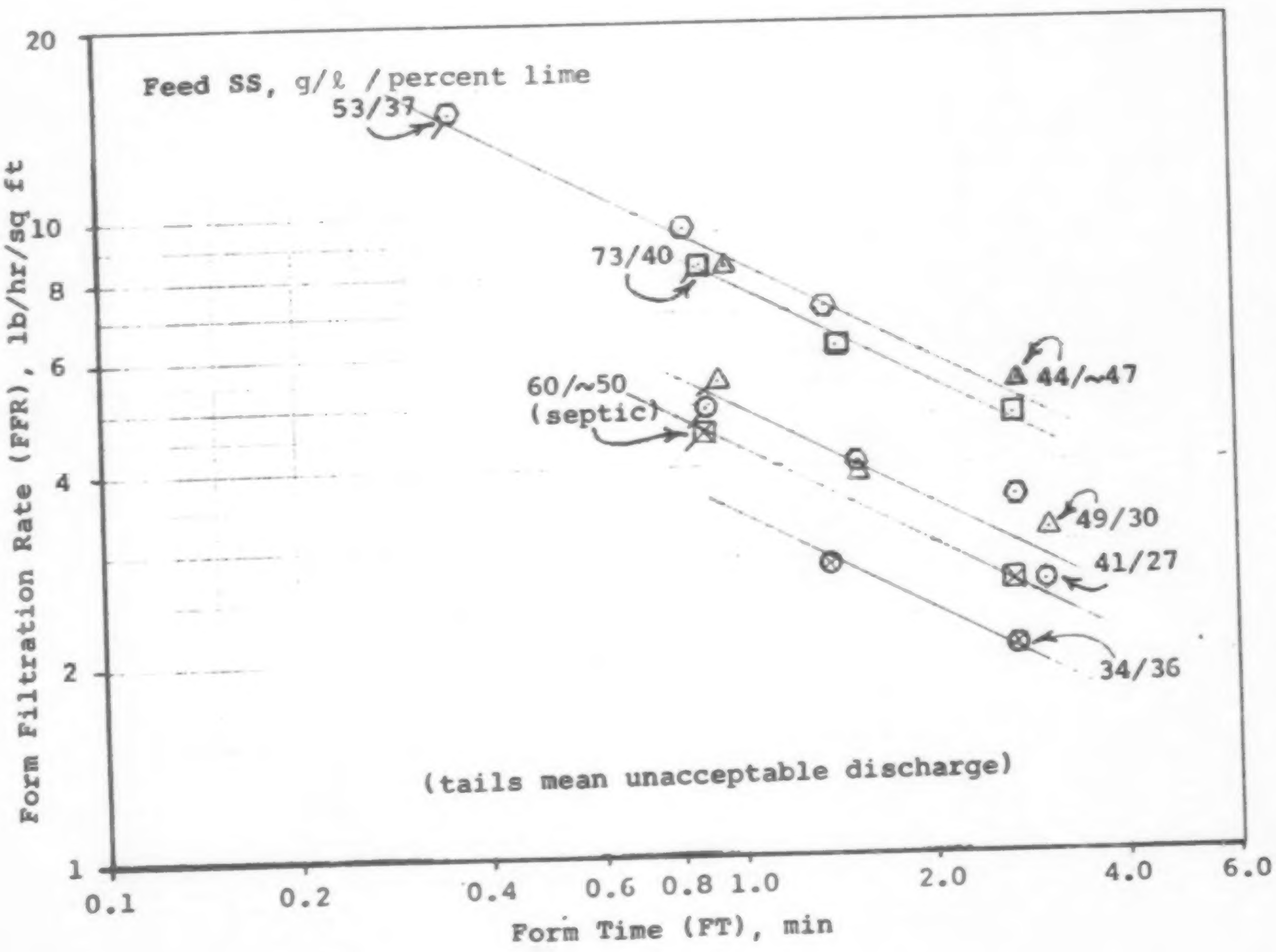


TABLE 7-17:

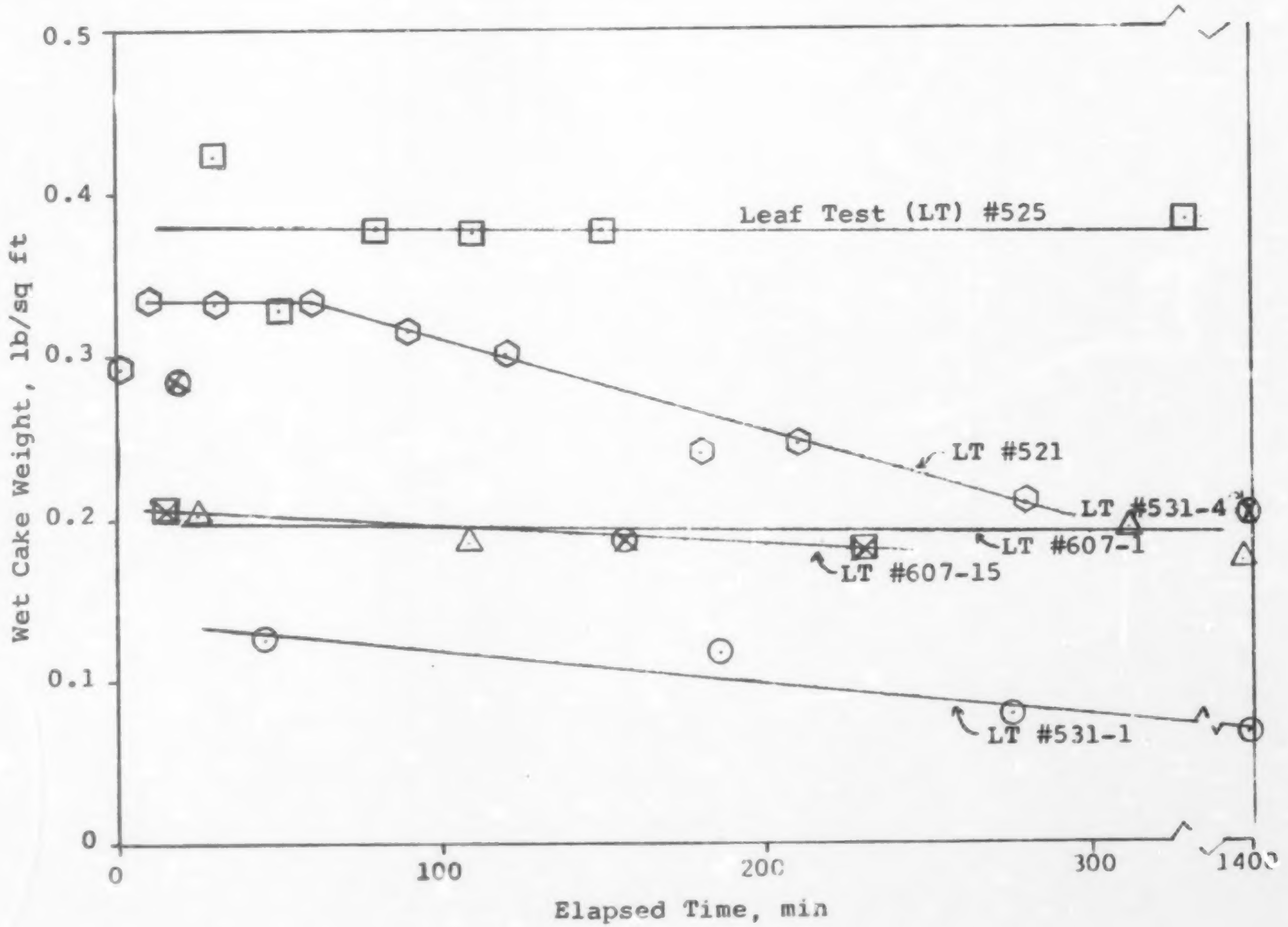
Alum-Primary Sludge Filtration

Ratio of Pilot Plant to Leaf Test Results

Run #	Form Filtration Rate	Filter Cake Moisture Content	Filter Cake Thickenss	Filter Cake Discharge[a]
419-1	1.62	1.04	2.0	1.0
419-2	1.47	1.04	2.0	1.5
419-3	--	--	1.5	6
419-4	0.41	0.99	0.36	0.7
419-5	--	--	1.5	--
426-1	0.86	0.99	0.67	1.5
426-2	0.50	0.99	0.50	0.7
426-3	0.65	1.01	1.0	1.0
427-1	0.59	1.00	0.62	0.6
427-2	0.79	1.01	0.83	0.7
427-3	0.74	1.01	1.0	1.0
427-4	1.14	--	1.0	1.8
516-1	0.86	1.00	1.0	1.5
516-2	1.02	1.01	0.75	1.5
516-3	2.01	0.99	2.01	2.0
525-1	0.99	1.01	1.0	--
525-2	0.74	0.99	1.0	--
531-1	1.12	1.00	1.0	--
531-2	0.83	1.00	1.5	--
531-3	1.06	0.96	0.67	--
531-4	0.66	1.01	0.25	--
531-5	1.11	1.00	1.16	--
604-1	0.52	1.10	0.5	--
604-2	0.46	1.06	0.33	--
604-3	0.59	--	0.5	--
607-1	0.91	1.01	1.0	0.8
607-2	0.91	1.00	0.5	1.1
607-3	1.09	0.99	1.33	2.6
607-4	0.45	1.04	0.33	0.2
611-1	0.85	1.01	1.0	0.3
611-2	0.85	1.03	1.0	1.0
611-3	0.92	1.01	1.0	1.2
419-611				
Average	0.89	1.01	0.96	1.4
Std. Dev.	0.35	0.03	0.47	1.2
531-611				
Average	0.82	1.02	0.80	1.1
Std. Dev.	0.23	0.03	0.39	0.8

[a] See Table 7-13 for discharge code

FIGURE 7-25: ALUM PRIMARY SLUDGE FILTRATION, LEAF TEST RESULTS - EFFECT OF ELAPSED TIME ON CAKE WEIGHT



but generally FFR decreases with time. Thus, many of the high ratios of pilot plant FFR to leaf test FFR shown in Table 7-17 might be due to a time lag effect. After Run #516, all simulation leaf tests were conducted on the same day as the pilot plant run and generally within 1/3 to 2 hours after sample procurement.

The only other reasonable explanation of the variable data in Table 7-17 would be inconsistent mixing of the lime slurry and sludge, and/or poor sampling. The latter is unlikely and the former almost impossible to evaluate after the fact.

The questionable lime mixing effectiveness prompted a brief study to compare pilot plant mixing with bench scale mixing procedures. This comparative study was conducted during two pilot plant vacuum filter runs as follows. During operation of the pilot plant lime mixer-flocculator, five different grab-composite sludge samples of flocculator feed (without lime added) and discharge (with lime added and mixed) were collected. Routine simulation leaf tests, at pilot plant operating conditions, were conducted on the flocculator discharge samples. About 1.5 liters of the flocculator feed samples were subjected to laboratory scale lime conditioning by adding the same dosage of lime as a 30 to 80 g per 100 to 200 mℓ of tap water paste. Mixing was accomplished with a large spatula. Simulation leaf tests were conducted on these five laboratory lime conditioned samples.

Flocculator feed samples for this comparative study contained from 41 to 49 g/ℓ of suspended solids. Lime dosages were 27 to 30 percent by weight and form times ranged from 0.9 to 5.5 minutes. The ratio of dry time to form time used was 1.5.

The average ratio for the five leaf test results on pilot plant flocculator discharge samples to laboratory scale lime conditioned flocculator feed samples were as follows:

Dry cake weight, W = 0.94
Cake moisture content, M_C = 0.98
Cake thickness = 0.96
Cake discharge = 1.03

It is apparent that near identical vacuum dewatering results were obtained when lime mixing-flocculation was accomplished by the continuous flow pilot plant unit or by closely controlled batch laboratory procedures. This observation would tend to negate the previous suggestion that the variability of results in Table 7-17 was due to inconsistent mixing of the lime slurry and sludge in the pilot plant unit.

The filter cake M_c comparison between pilot plant and leaf test results was fairly precise and indicated near identical results. Figure 7-26 shows M_c results for high and low feed solids and lime dosage data as a function of the simplified correlating factor. The data indicates that high feed solids and lime dosage results in slightly lower M_c compared to low lime dosage. For low lime dosages, the effect of alum-primary feed solids concentration of M_c is nominal. Figure 7-27 shows filter cake M_c versus θ_D/W for moderate lime dosages of 30 to 44 percent. The data are seen to generally range between 72 to 76 percent M_c and are not significantly affected by alum-primary feed solids concentrations.

Because pilot plant results did not show a definable trend in the effect of feed solids concentration and percent lime conditioning on FFR, several leaf tests were conducted in an attempt to do so. Figure 7-28 shows the results of these tests. Note that filter yield and not FFR is shown and that the yield is for the operating conditions shown. The wavy lines indicate where acceptable cake discharge was experienced. As previously noted, lime dosages greater than 20 percent were required to produce dischargeable cakes.

The data in Figure 7-28 generally indicate that above some lime dosage level, a very rapid increase in yield is experienced for an incremental increase in lime dosage. For example, the steep part of the curves show that for an incremental increase of 10 percent lime (e.g., from 20 to 30 or 30 to 40 percent), a 2 to 3 fold increase in yield was experienced. This observation suggests that an economic optimum might exist. It would appear that facilities should be provided for feeding 40 to 50 percent lime, but that actual operation may require as little as 25 to 30 percent lime on certain occasions.

Other Studies

Pilot plant and leaf test data were evaluated in an attempt to identify a general relationship between the level of lime addition and the increase in sludge solids. Pilot plant data indicated an average of about 0.6 g increase in solids for each 1.0 g of lime added (0.3 to 0.9 range). Data from several laboratory scale tests indicated a 1.0 g increase in solids for each g of lime added. Since the FFR's were reported on the basis of dry alum-primary solids, the ratio of feed solids before and after lime addition had to be used in computing FFR's. The observed variability in suspended solids increase with addition of lime may have imposed an added variability in the FFR results, thus precluding definition of the precise effects of feed solids and lime

FIGURE 7-26: ALUM PRIMARY SLUDGE FILTRATION - EFFECT OF FEED SUSPENDED SOLIDS AND LIME ON CAKE MOISTURE CONTENT

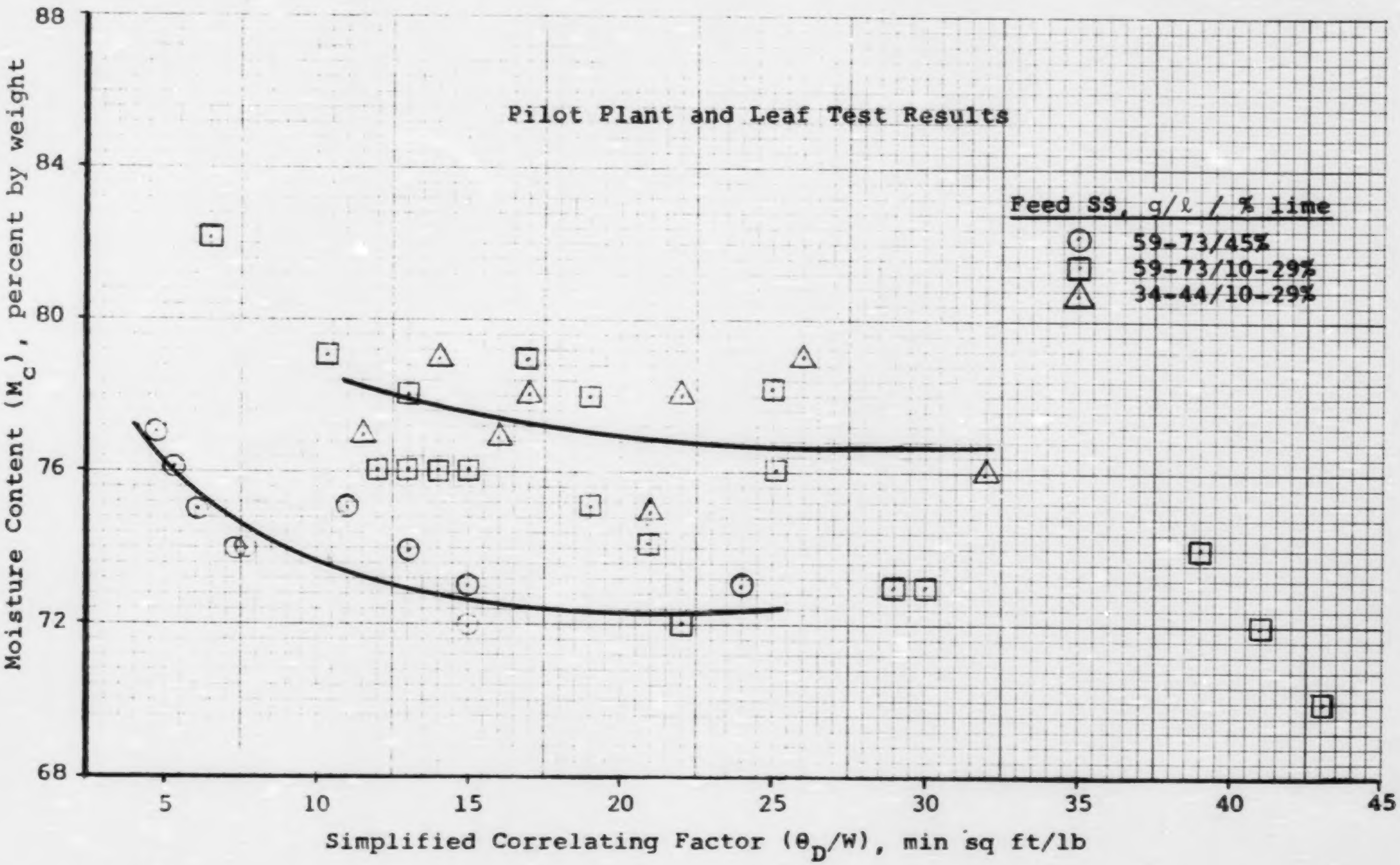


FIGURE 7-27: ALUM PRIMARY SLUDGE FILTRATION - CAKE MOISTURE CONTENT

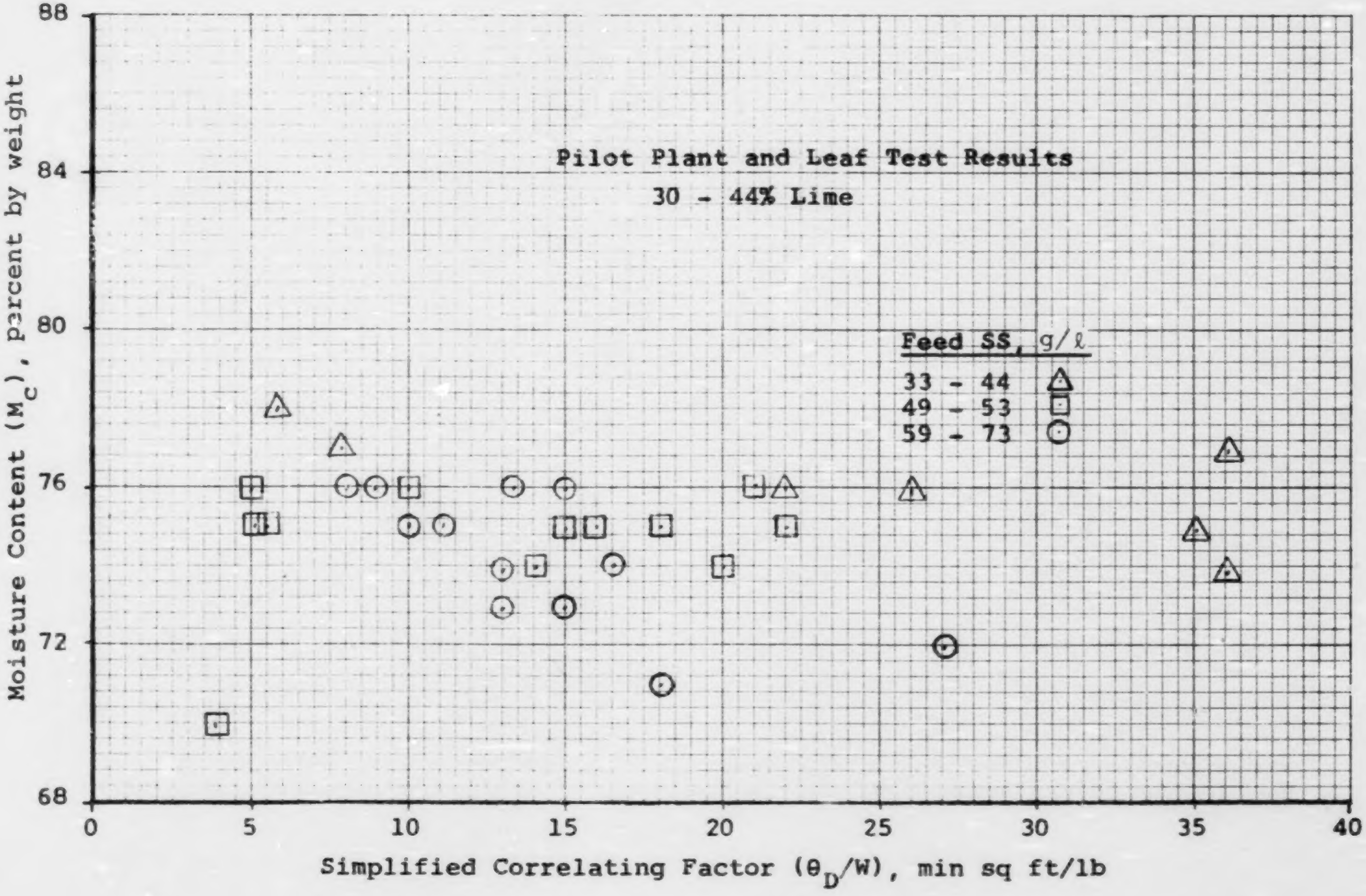
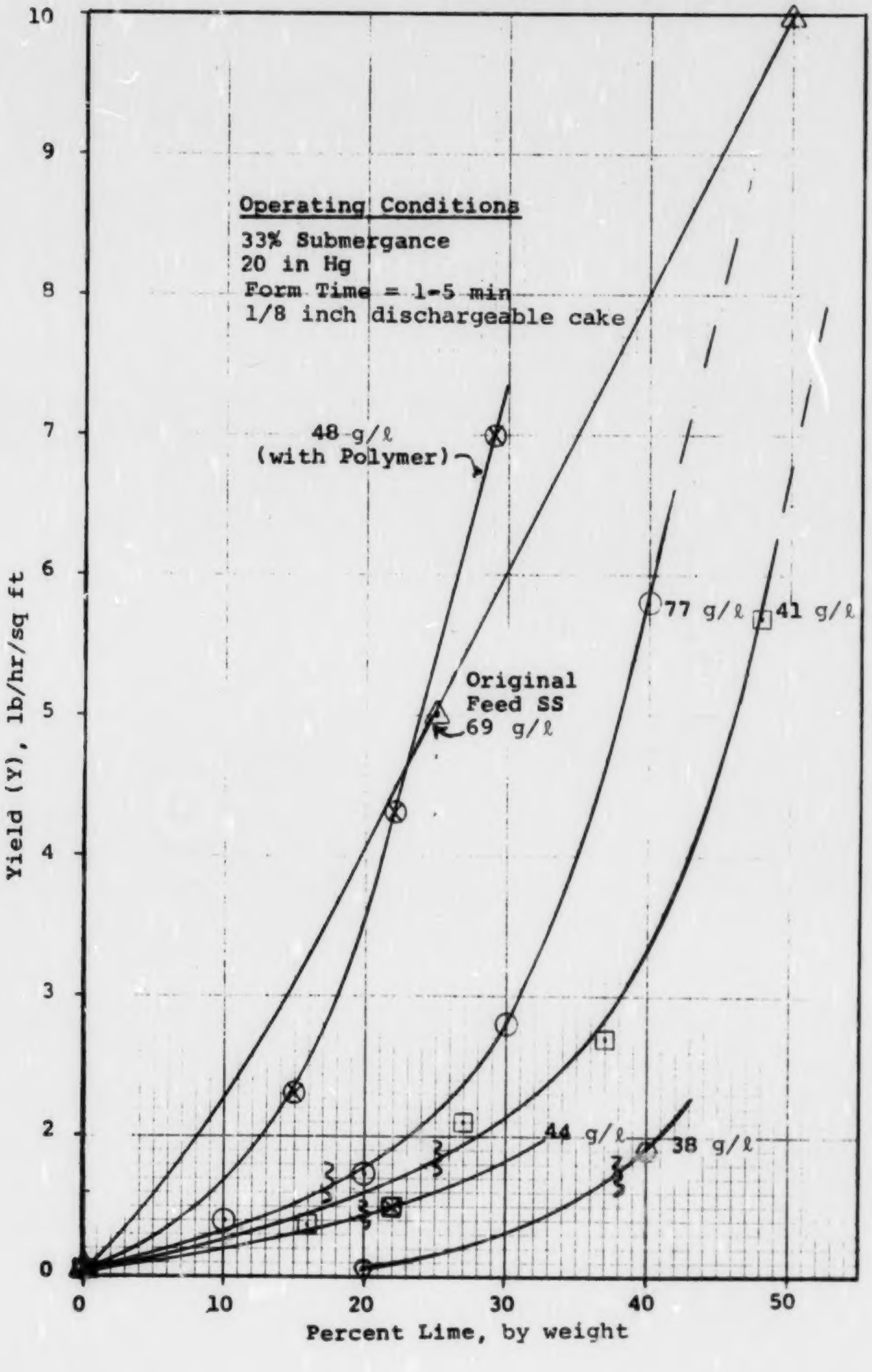


FIGURE 7-28: ALUM PRIMARY SLUDGE FILTRATION, LEAF TEST RESULTS - EFFECT OF LIME DOSAGE



on performance parameters.

Note that the vacuum filter feed sludge pH's indicated in Table 7-16 are in excess of 9. Alum sludge, resulting from coagulation reactions, is an amphoteric material which is dissolved at pH's in excess of 9. This condition is substantiated by data presented in Figure 7-29 which shows the grams of soluble aluminum present in an alum-primary sludge sample at various solids concentrations as a function of slurry pH. Essentially all alum present was dissolved at a pH of 11.7. At pH's of 10.7 to 11.0, the amount of aluminum dissolved was inversely proportional to the solids concentration. In practice, soluble aluminum would report to the vacuum filter filtrate and recycled to the chemical treatment unit. Presumably this recycled soluble aluminum could act as an adjunct to the alum added. With lime conditioning, any phosphorus released upon aluminum solubilization would be precipitated as a calcium-phosphorus compound. Though a phosphorus mass balance around the vacuum filter was not made, it is presumed that most calcium-phosphorus precipitate reported to the filter cake. It would be important to design and operate the vacuum filter to achieve high solids captures to minimize recycling phosphorus material to the chemical treatment unit.

Summary Discussion

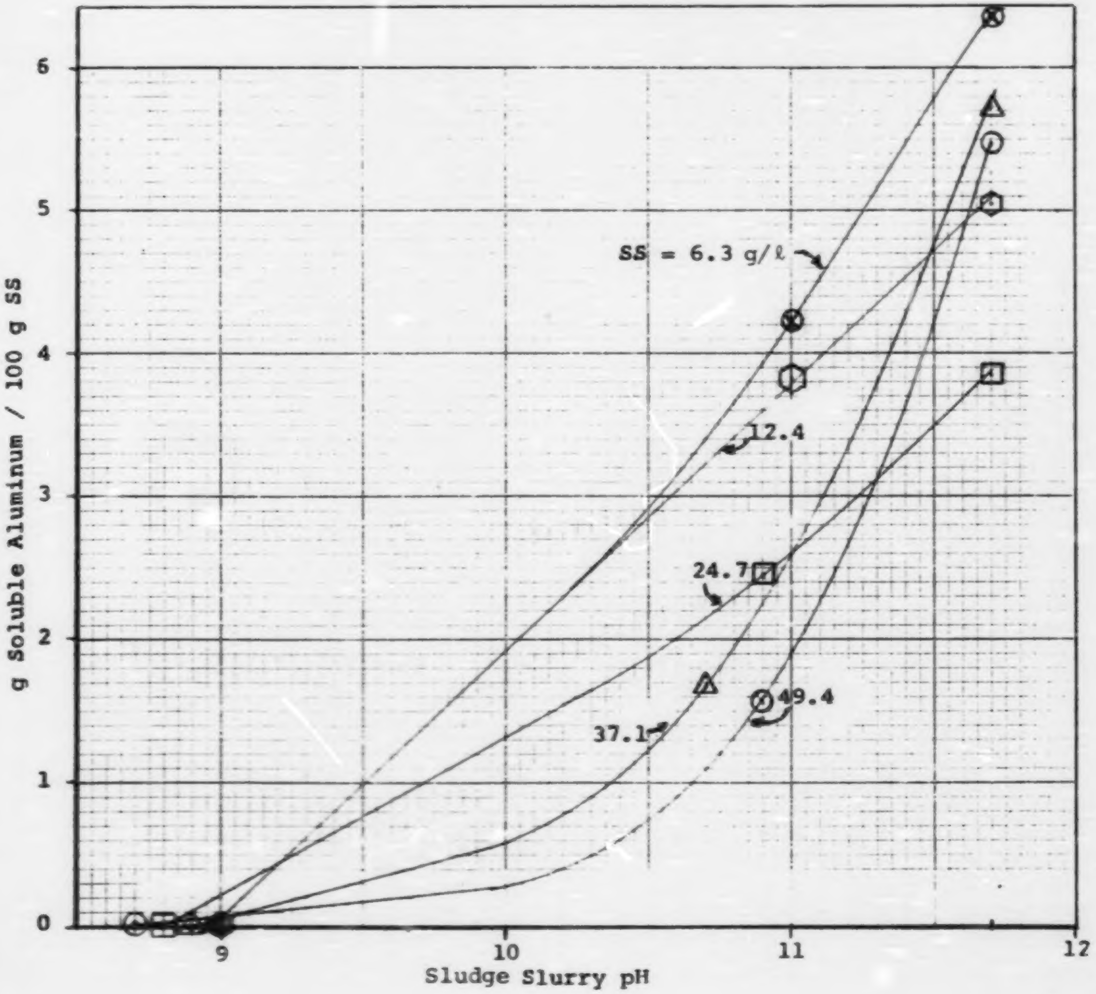
Vaccum filter cakes of 1/8 inch discharged easily from the nylon media (NY-317F) with lime conditioning at dosages in excess of about 20 percent by weight. Nylon media would be recommended even though it is fragile and alignment difficulties may occur on large machines.

All pilot plant and leaf test data were used to determine an average filter cake dry solids density. A value of 21 lb/cu ft was determined.

A precise designation of lime dosage required to produce a finite filter yield is impossible due to variations in feed solids concentrations and characteristics (e.g., septicity, percent alum sludge, etc.). For applications similar to this study, lime feeding facilities should be provided for 50 percent by weight lime dosage.

In the authors' judgement, the data presented indicated that for lime dosages of 30 to 40 percent by weight and feed solids concentrations in excess of 40 g/ℓ, that a full scale filter

FIGURE 7-29: ALUM PRIMARY SLUDGE, LABORATORY RESULTS - EFFECT OF SLUDGE pH ON ALUMINUM SOLUBILITY



yield of 2 lb of dry alum-primary solids/sq ft/hr could be achieved. A filter cake M_c of about 75 percent, based on total filter cake solids, would be expected.

Based on the filter vat mixing problems encountered in this study, care should be exercised in providing good mixing conditions. Lime conditioning of the alum-primary sludge prior to gravity thickening might be worth considering. The effect of lime addition prior to gravity thickening on TSL and performance would have to be evaluated.

Ferric-Primary Sludge Filtration

In Section VI and earlier in this Section (Ferric-Primary Sludge Thickening), it was suggested that the anaerobic conditions existing in the chemical treatment unit resulted in precipitation of FeS. The FeS precipitate did not adversely effect the performance of the chemical treatment unit or the gravity thickener, but its presence had a devastating effect on vacuum filtration dewatering.

Of six pilot plant runs attempted, three were aborted due to an inability to produce a dischargeable filter cake. Several leaf test series were attempted in an effort to determine suitable chemical conditioning to produce a dischargeable cake.

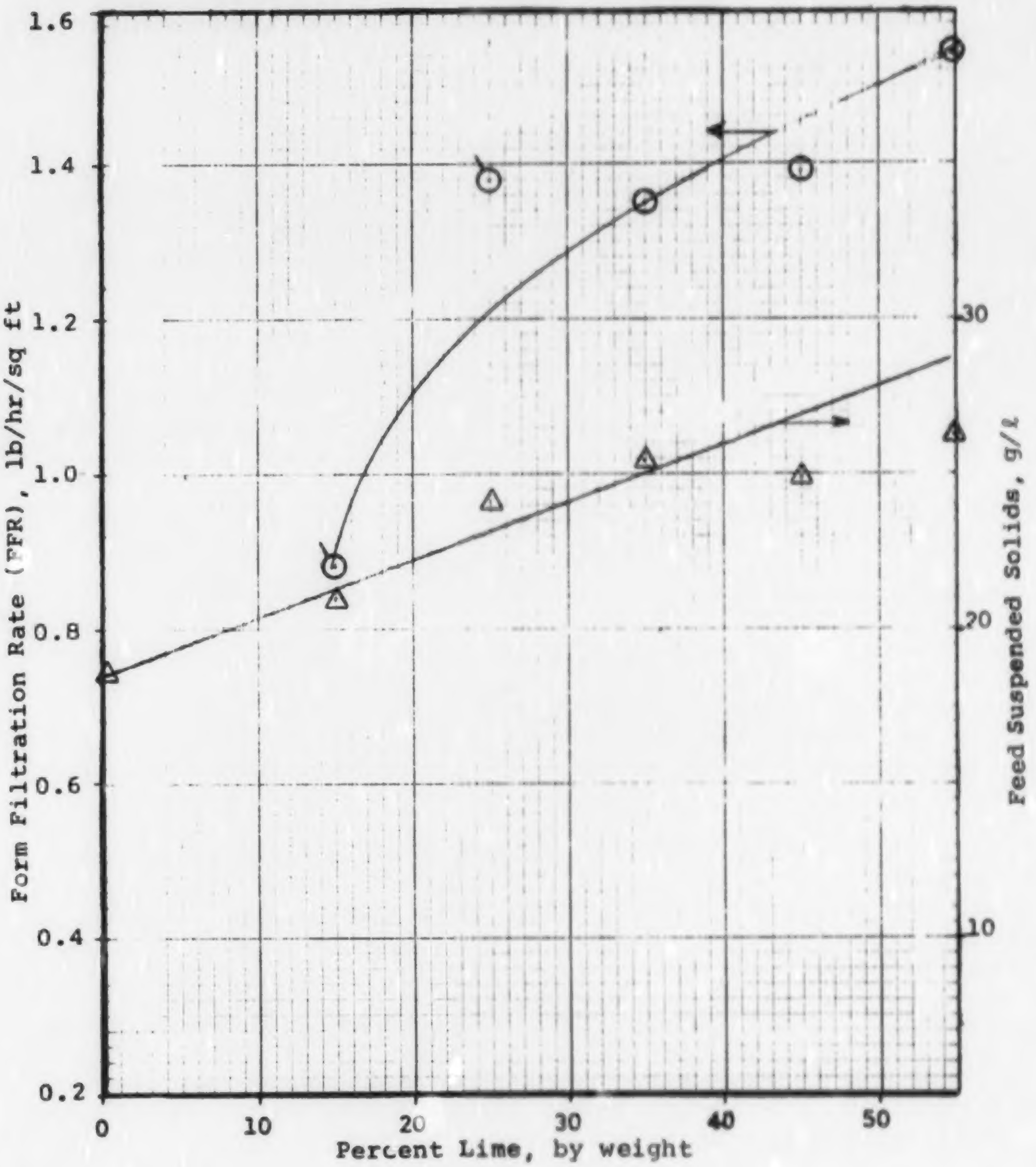
Preliminary Leaf Tests

Preliminary leaf tests indicated a cake discharge problem. Screening of several media, mostly nylon types, failed to indicate any media with discharge properties superior to those of NY-317F medium. NY-317F was used for alum-primary sludge vacuum filtration and thus a new medium was not required.

Figure 7-30 shows a typical response of FFR versus percent lime added. Lime in excess of 30 percent was required to produce a marginally acceptable cake discharge. Because of marginal leaf test performance results, problems with pilot plant operation was expected.

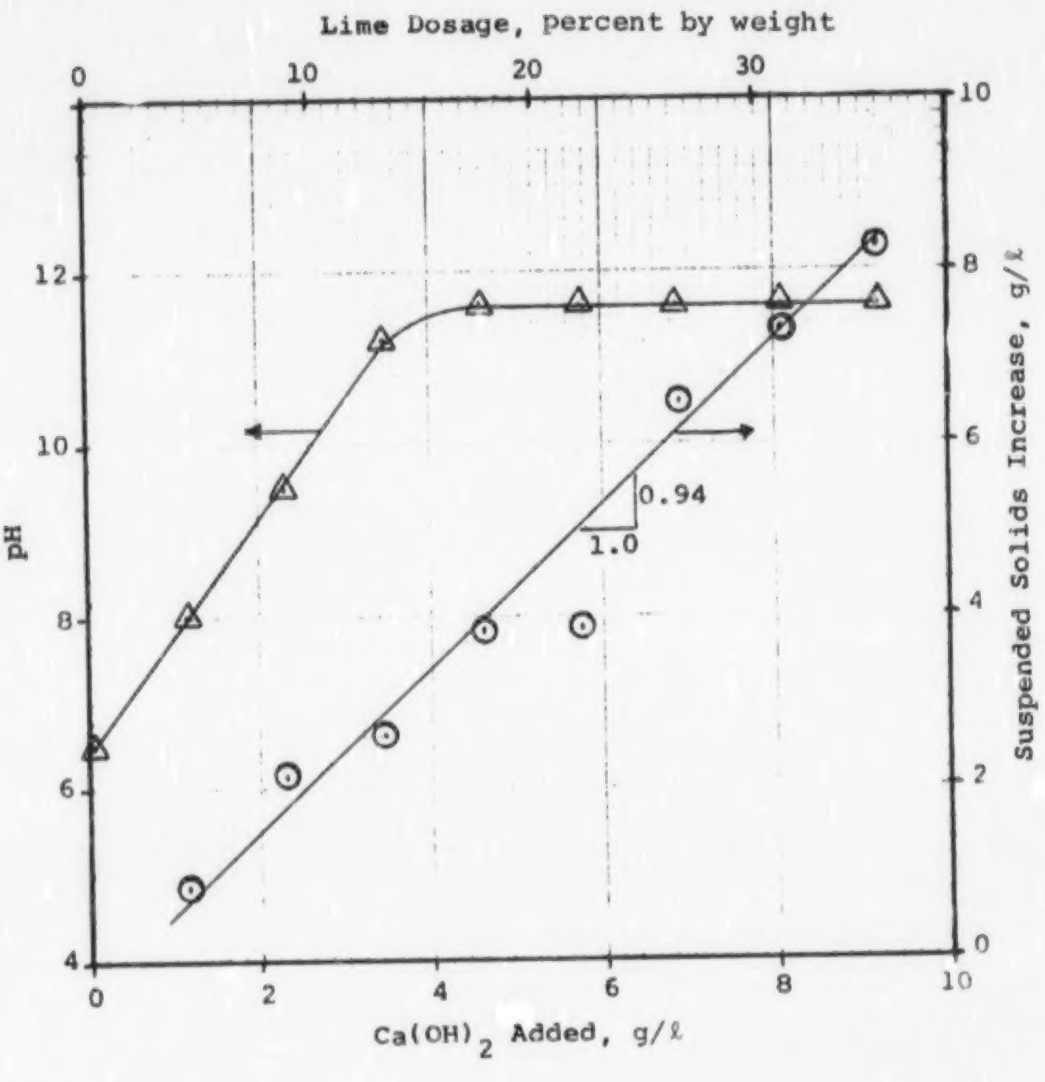
A controlled bench scale test was conducted to show the pH and solids increase responses due to lime addition to the ferric-primary sludge. Figure 7-31 shows that the obvious effect of lime addition is to increase pH and uniformly increase feed solids concentration.

FIGURE 7-30: FERRIC PRIMARY SLUDGE FILTRATION, LEAF TEST RESULTS - EFFECT OF LIME DOSAGE



Note: Tails mean poor Discharge.

FIGURE 7-31: FERRIC PRIMARY SLUDGE - EFFECT OF LIME DOSAGE



Pilot Plant Results

Table 7-18 presents the limited pilot plant data obtained. It is readily apparent that low FFR's and thin cakes were the rule. Though the data indicates that very thin cakes were found to be dischargeable, it is not normal practice to design full scale units to continuously discharge such thin cakes.

The relatively low FFR's obtained were probably partly due to the moderate to low levels of lime dosage used. A plot of the data in Table 7-18 as FFR versus FT indicated very inconsistent relationships.

About midway through these tests, the filter media belt was cleaned with dilute acid. It was hoped that previous use with lime conditioning might have caused clogging or blinding of the media which could be eliminated by an acid wash. No perceptable effect on performance was found.

Leaf Test Results

Because of the generally poor results and limited pilot plant runs, limited leaf test simulations were conducted. These tests were done only for pilot plant Runs #726 and #801. For eight comparisons, the ratio of pilot plant to leaf test FFR ranged from 0.97 to 1.70, averaging 1.40. The M_C ratios were much more consistent and averaged 0.97.

In addition to three aborted pilot plant runs, on four other occasions leaf tests were made in an attempt to obtain a dischargeable filter cake at lime dosages less than 50 percent by weight. No success was experienced. It had become apparent that the presence of finely divided precipitate (FeS) precluded producing a dischargeable cake when using only lime as the conditioning chemical.

Limited leaf tests were conducted in an attempt to identify a type and dosage of polyelectrolyte which would flocculate the troublesome precipitate. These results were negative.

The use of a primary coagulant, $FeCl_3$, was finally tried as an adjunct to the lime conditioning chemical. Acceptable discharge and FFR's were obtained using the combination of 40 percent lime and 10 percent $FeCl_3$.

Based on these results, the last pilot plant Run, #912, was made using both chemicals. The results in Table 7-18 show that acceptable performance was achieved.

TABLE 7-18: FERRIC-PRIMARY SLUDGE: VACUUM FILTRATION PILOT PLANT RESULTS

Run #	726-1	726-2	726-3	726-4	730-1	801-1	801-2	801-3
Operating Conditions:								
Form Time(FT), min[a]	2.9	2.1	1.7	1.27	3.05	3.00	2.10	1.54
Dry Time(θ_D), min	4.1	2.9	2.16	1.33	4.29	4.2	2.9	2.2
Vacuum, in. of Hg	20	20	21	21	20	20	20	21
Lime Dose, % by wgt	43	46	49	53	25	--	--	--
Sludge, pH	10.3	11.2	11.2	11.3	11.9	11.9	11.9	11.9
Solids:								
Feed, g/ℓ	29.4	29.4	29.4	29.4	48.2	49.3	49.3	49.3
Feed & Lime, g/ℓ	34.8	34.8	34.8	34.8	53.6	49.4	49.4	49.4
Filtrate, g/ℓ	0.78	0.94	4.5	2.7	0.9	2.2	8.6	0.36
Filter Cake:								
Thickness, in.	1/8	3/16	3/32	1/16	1/32+	3/32	1/16+	1/32
Discharge Code[b]	E	E	E	G	E	E	F	P
Dry Weight(W), lb/sq ft	0.183	0.175	0.148	0.115	0.135	0.165	0.112	0.0809
θ_D/W, min sq ft/lb	22	17	15	12	32	25	26	27
Performance:								
Form Filtration Rate (FFR) lb/hr/sq ft	3.20	4.22	4.41	4.59	2.39	2.74	2.85	2.61
Cake Moisture, %	66	69	67	68	68+	69	70	70
Solids Capture, %	97	97	85	91	98	95+	83	99+

[a] FT/CT = 0.28

[b] See Table 7-13

TABLE 7-18 (CONT.): FERRIC-PRIMARY SLUDGE: VACUUM FILTRATION PILOT PLANT RESULTS

Run #	801-4	808-1	820-1	912-1	912-2	912-3
Operating Conditions:						
Form Time(FT), min[a]	1.19	3.01	2.72	3.30	1.56	0.97
Dry Time(θ_D), min	1.7	4.23	3.81	4.63	2.19	1.36
Vacuum, in. of Hg	23	21	20	22	22	22
Lime Dose, % by wgt	--	--	30	28	28	28
$FeCl_3$ Dose, % by wgt	--	--	--	12	12	12
Sludge, pH	11.9	--	--	8.4	8.4	8.4
Solids:						
Feed, g/ℓ	49.3	28.1	64.3	57.3	57.3	57.3
Feed & Lime, g/ℓ	49.4	33.3	64.8	68.8	68.8	68.8
Filtrate, g/ℓ	3.0	2.3	0.62	0.17	0.28	--
Filter Cake:						
Thickness, in.	--	1/32-	nil	1/8	3/32	3/32
Discharge Code[b]	X	F	X	E	P	E
Dry Weight(W), lb/sq ft	0.0557	0.0425	0.0302	0.193	0.116	0.166
θ_D/W, min sq ft/lb	31	--	12	24	19	8.2
Performance:						
Form Filtration Rate (FFR) lb/hr/sq ft	2.33	0.72	0.5	2.92	3.72	8.55
Cake Moisture, %	74	70	73	72	75	73
Solids Capture, %	94	--	99	99+	99+	--

[a] FT/CT = 0.28

[b] See Table 7-13

Summary Discussion

The presence of a finely divided black precipitate in the ferric-primary sludge precluded achieving effective filtration with lime conditioning. A combination of lime and $FeCl_3$ conditioning was found to provide acceptable results, but only at a predicted chemical cost of about $20/ton of dry solids.

Spent Carbon Sludge Filtration

Vacuum filtration dewatering of thickened spent carbon was done primarily to produce regeneration furnace feed. Previous experience had indicated solids capture was poor unless cationic polyelectrolyte was used.[1] It was also conjectured that the filter media used was too open and that a tighter media should be used. The media suggested was polypropolene-873 (POPR-873). This media was used during this study.

Thirteen filter runs were conducted. Eleven of these runs were in conjunction with carbon regeneration.

Five series of leaf tests were conducted simulating pilot plant operating conditions. Other leaf tests were conducted to obtain background information on certain operational variables.

Pilot Plant Results

Pilot plant operating and performance data are presented in Table 7-19. Data for most runs are averages for several data collection series. For example, during Run #72872, the vacuum filter was operated and data collected three times with similar feed and operating conditions. The performance results shown in Table 7-19 for Run #72872 are an average of these three operating times.

As noted in Table 7-19, when the filter cake thickness was less than 3/8 inch, less than "good" cake discharge was experienced. For filter cakes less than 1/2 inch thick, use of the "doctor blade" was judged to be necessary.

Solids capture results indicate a gradual decrease from the low nineties to about 80 percent over about a year.

Without exception, leaf test results always indicated solids captures of greater than 99 percent. During pilot plant operation, substantial deposits of carbon solids were routinely observed at the drum headbolt-heads and at the

TABLE 7-19: CARBON SLUDGE: VACUUM FILTRATION PILOT PLANT RESULTS

Run #	72872	80172	81672	81672	91272	91272	92872	92872	122 872-1	122 872-2
Operating Conditions:										
Form Time(FT),[a] min	1.35	0.64	0.67	0.67	0.75	1.35	0.85	0.95	0.40	1.09
Dry Time,(θ_D),min	1.87	0.89	0.90	0.90	1.05	1.88	0.27	0.30	0.128	0.349
Vacuum, in.of Hg	5	6	6	6	11	11	11	12	9-12	9-12
# of Data Points	3	2	6	1	9	3	4	3	1	1
Solids:										
Feed, g/ℓ	94	173	104	147	136	152	122	124	154	154
Filtrate, g/ℓ	5.4	8.4	10.5	--	21	29	22	29	16	12
Filter Cake:										
Thickness, in	1/2-	1/2	5/16	3/8	1/2	7/8	1/2+	1/2+	1/4	3/8
Discharge Code[b]	G	G	G	G+	G	G	G	G	G	G
Dry Weight(W), lb/sq ft	--	--	--	--	--	--	--	--	0.384	0.644
θ_D/W,min sq ft/lb	34	65	47	60	53	52	47	43	0.33	0.54
Performance:										
Form Filtration Rate (FFR)lb/hr/sq ft	28-	56	40	51	54	53	48	45	57	36
Cake Moisture, %	77	77-	77-	75	77+	76	78	78	--	--
Solids Capture, %	94	95	90	--	85	81	82	77	90	92

[a] FT/CT = 0.28

[b] See Table 7-13

TABLE 7-19 (CONT.): CARBON SLUDGE: VACUUM FILTRATION PILOT PLANT RESULTS

Run #	122 872-3	010 373-1	010 373-2	010 373-3	40473	50573	61573	71 973-1	71 973-2
Operating Conditions:									
Form Time(FT), ^{a}min	3.04	0.50	0.86	2.99	0.67	1.4	2.8	2.8	2.8
Dry Time(θ_D),min	0.96	0.16	0.275	0.957	1.01	2.1	4.2	4.06	4.06
Vacuum, in.of Hg	9-12	12-15	11-16	10-15	7+	10	5	4	11
Chem. Dose^b lb/TDS	--	--	--	--	--	--	6.5	9.2	8.4
# of Data Points	1	1	1	1	4	3	5	1	1
Solids:									
Feed, g/ℓ	154	160	161	160	160	134	120	(120)	(120)
Filtrate. g/ℓ	17	31	23	11	30	41	15	--	--
Filter Cake:									
Thickness, in.	1	1/4	3/8	3/4	1/4-	7/32	1/2	3/4	9/8
Discharge Code	G	F	G	G+	F	F	E	--	--
Dry Weight(W), lb/sq ft	0.937	0.347	0.459	0.840	--	--	--	1.054	1.498
θ_D/W, min sq ft/lb	1.0	0.46	0.60	1.1	33	13	19	3.9	2.7
Performance:									
Form Filtration Rate (FFR) lb/hr/sq ft	19-	41	32-	17	30	13	15	23	32
Cake Moisture, %	--	78	78	78	75	74+	76-	76+	77
Solids Capture, %	89	81	86	93	81	70	88	--	--

[a] Starting with Run #40473, FT/CT = 0.26

[b] Cationic Polyelectrolyte

TABLE 7-19 (CONT.): CARBON SLUDGE: VACUUM FILTRATION PILOT PLANT RESULTS

Run #	71 973-3	71 973-4	71 973-5	71 973-6	91473	100873
Operating Conditions:						
Form Time(FT), min	1.4	2.8	0.70	2.8	2.8	2.8
Dry Time(θ_D), min	2.03	4.06	1.02	4.06	4.2	0.84
Vacuum, in.of Hg	15	4	6	5	12	15
Chem. Dose, lb/TDS	6.4	4.9	0	4.0	--	--
# pf Data Points	1	1	1	1	3	4
Solids:						
Feed, g/ℓ	(120)	122	120	121	160	170
Filtrate, g/ℓ	--	2.3	12	16	--	--
Filter Cake:						
Thickness, in.	5/8	1/2	1/4	5/8+	1/2-3/8	3/4
Discharge Code	--	E	F	G	--	--
Dry Weight(W), lb/sq ft	0.824	0.707	0.393	0.475	--	1.3
θ_D/W, min sq ft/lb	2.5	5.7	2.6	8.5	11	12
Performance:						
Form Filtration Rate (FFR) lb/hr/sq ft	35	15	34	10	12	14
Cake Moisture, %	77	75	75	75	--	73
Solids Capture, %	--	--	--	--	--	--

belt-drum edge location. These depositions indicated substantial leakage at these locations. Once a deposition was built up, no leakage of solids would be expected. Thus, the very poor captures experienced during some runs may have been due to some other cause(s). The only consistent relationship of solids capture and a machine operation variable found is indicated in Figure 7-32. The data show that at very low FT's (i.e., high drum rotational speed), very low solids captures resulted. Had dry cake weight (W) been plotted instead of FT, a more linear response would have been indicated. The response shown in Figure 7-32 was not found to exist for all results, thus it should not be interpreted as a general observation.

It is possible that the poor solids captures experienced was due to the type of filter media used. Leakage of solids through a monofilament can occur, particularly when no vacuum is applied and the section of the filter drum covered with clean media is entering the slurry. This type of phenomenon is consistent with the results shown in Figure 7-32. At lower form times or high drum rotational speeds, more opportunitites for "leakage" of solids would occur per unit time or quantity of sludge processed.

Evaluation of all available M_C results indicated quite variable data, but a general trend toward higher M_C of low θ_D/W's was evident. The data was "smoothed" by grouping θ_D/W values and averaging M_C values. The results are shown in Figure 7-33. The "smoothed" data shows a decreasing trend in M_C up to a θ_D/W value of about 5 minutes sq ft/lb. Also shown in Figure 7-33 are results for a filter run using virgin carbon. The spent (once used) carbon, with its associated organics and chemical floc, dewatered to lower M_C's than virgin carbon.

As indicated above, a pilot plant filter run (#102073) was conducted using virgin carbon. The results of this run are shown in Figure 7-34. During this run, a vacuum of 6 inches of Hg was used and the ratio of θ_D/FT was 1.5. The 1/4 inch filter cakes produced at FT's less than 0.5 minutes were judged to exhibit "good" discharge properties.

Scrutiny of all available pilot plant results (summarized in Table 7-19) indicated that a wide variation in vacuum levels were experienced. The main reason for the variation in vacuum levels and the generally low levels experienced was due to the nature of the dewatered carbon filter cake and the type of machine used. The filter cake routinely exhibited gross cracking due to shrinkage. Use of very short θ_D's, of 30 percent of FT or 8 percent of CT, reduced

FIGURE 7-32: CARBON SLUDGE FILTRATION, PILOT PLANT RESULTS - EFFECT OF FORM TIME ON SOLIDS CAPTURE

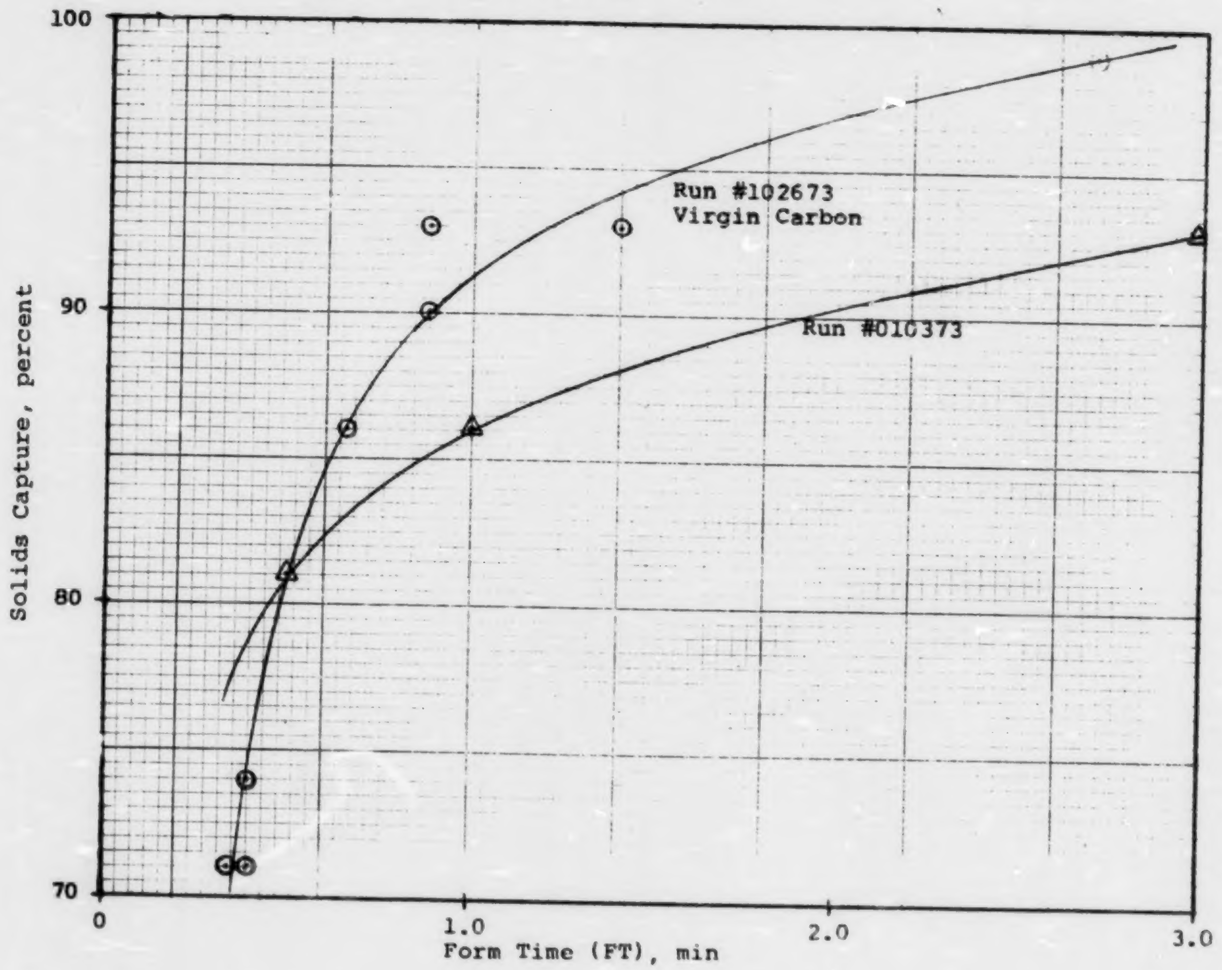


FIGURE 7-33: CARBON SLUDGE FILTRATION, PILOT PLANT RESULTS - CAKE MOISTURE CONTENT

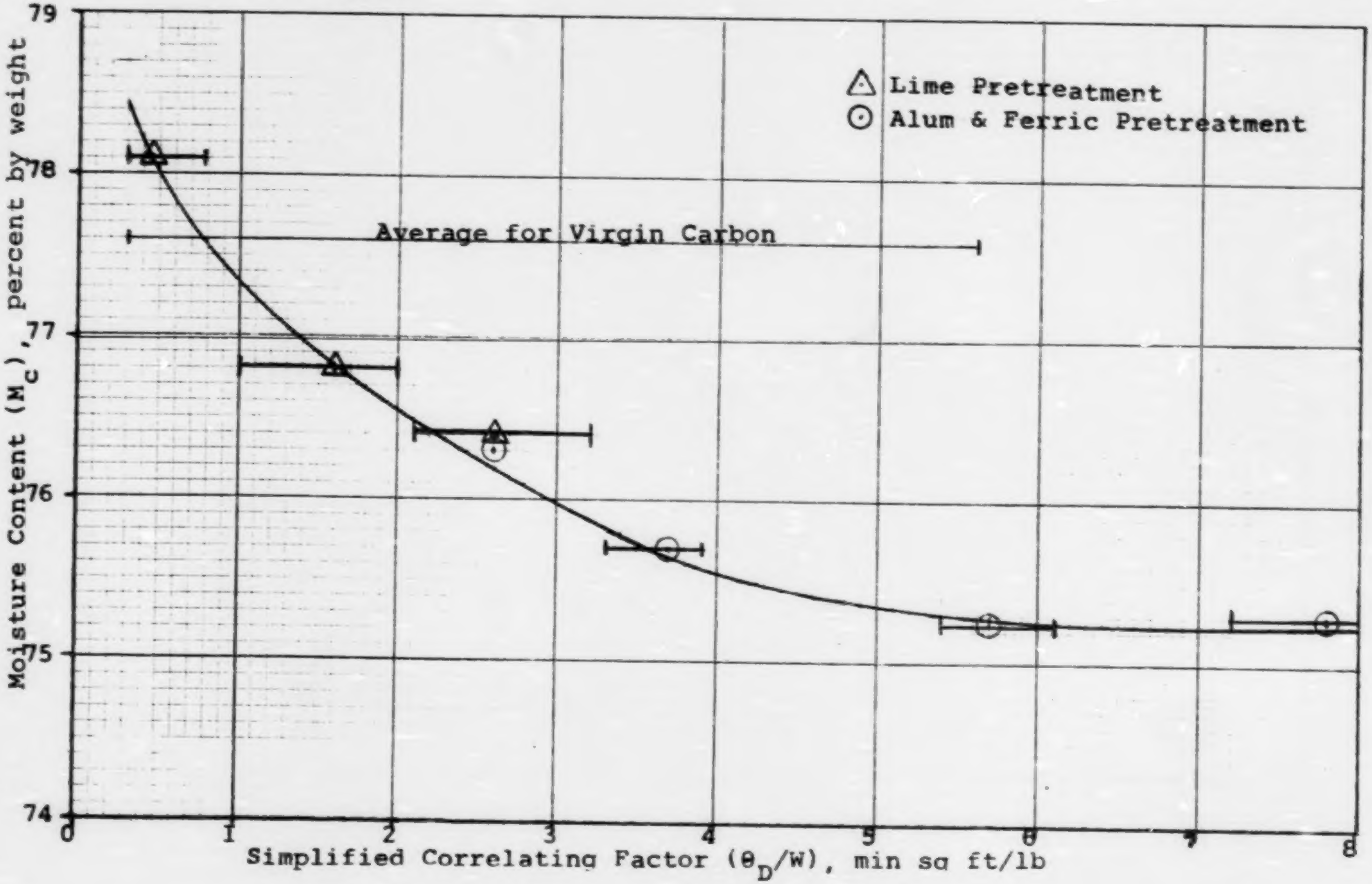
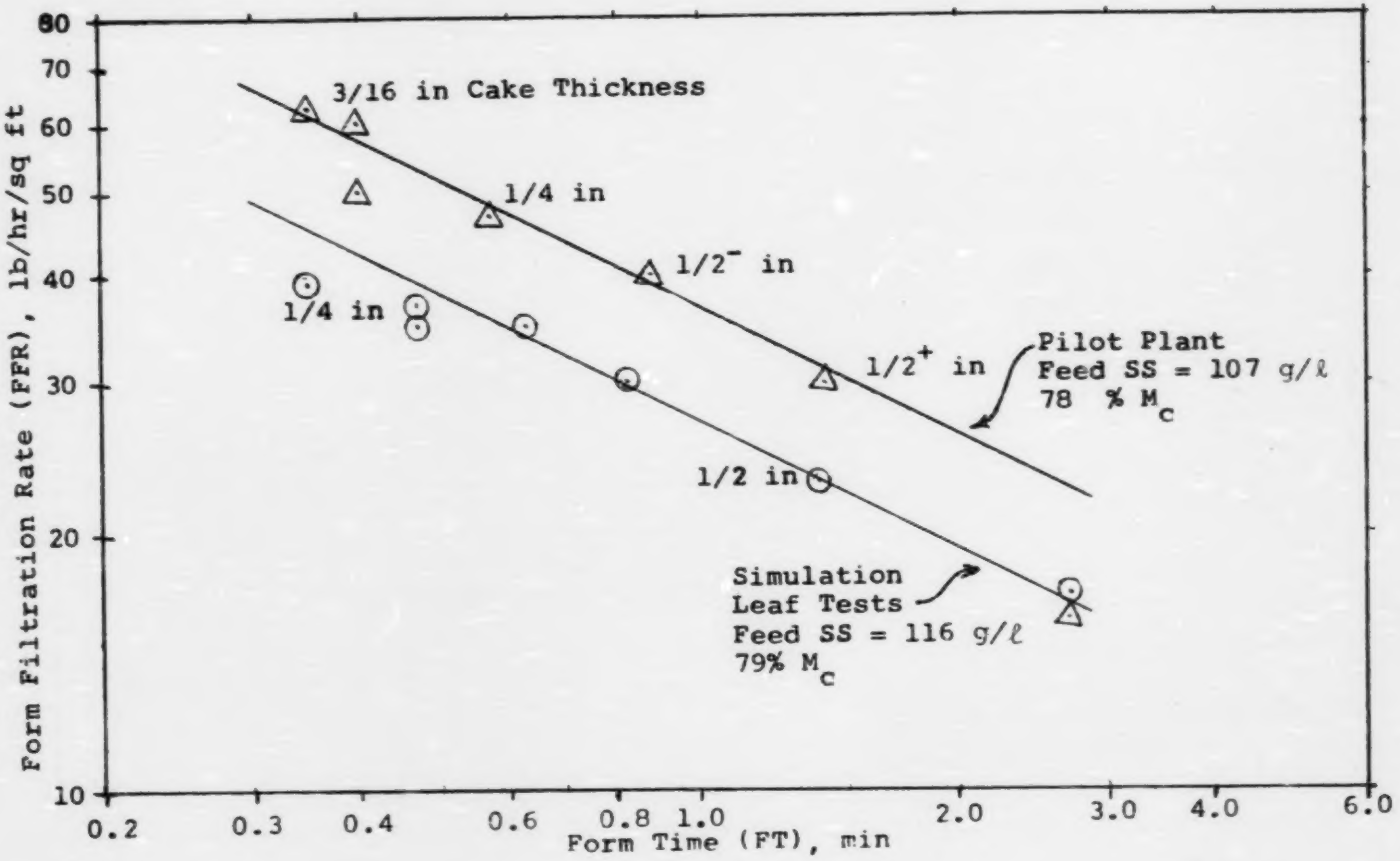


FIGURE 7-34: CARBON SLUDGE FILTRATION - VIRGIN CARBON RESULTS



cake shrinkage and thus cracking. However, M_c was high (see Table 7-19 and Figure 7-33). Due to the subsequent regeneration furnace operation, M_c should be minimized. On one occasion an attempt to reduce cake cracking was made using a "drag blanket." The "drag blanket" was a piece of gunny sack (burlap) which was kept wet. No observable reduction in cake cracking tendencies was observed.

The vacuum filter station used had only one vacuum receiver for both the form and dry cycles. Consequently, when the filter cake cracked during the dry cycle the vacuum level was reduced, not only for drying but for cake forming. Obviously separate form and dry vacuum receivers would have been desirable.

Three series of leaf tests were conducted to determine the effect of form vacuum level on FFR. The results of these tests are shown in Figure 7-35. For a non-compressible cake the slopes should have theoretically been 0.50. The lower values indicated in Figure 7-35 indicate that the spent carbon filter cakes were compressible. It would appear that a conservative estimate of the slopes shown would be 0.3. Thus the relationship between FFR and vacuum level (ΔP) would be:

$$FFR = K\ (\Delta P)^{0.3} \qquad (1)$$

For a given sludge, K remains constant and thus equation (1) can be used to predict FFR at various levels of form vacuum, if FFR is known at a given vacuum level.

No consistent relationship between feed solids and FFR is apparent from data presented in Table 7-19. If however the effect of varying form vacuum levels is factored out, a relationship can be shown for results obtained when lime was used in the chemical pretreatment step. Figure 7-36 shows predicted FFR at several feed solids concentrations for a form vacuum of 10 inches of Hg. The FFR values in Table 7-19 were adjusted using equation (1). The line shown in Figure 7-36 indicates that, if 33 percent effective submergence is assumed, an increase in filter yield of one lb/hr/sq ft would be expected for each increase of approximately one percent in feed solids concentration.

Also shown in Figure 7-36 is a data point for virgin carbon Run #102073 ☆ . It is apparent that use of virgin carbon in the pilot plant resulted in no deterioration of its dewatering rate. This observation is understandable when considering that the once spent carbon properties were very similar to those of virgin carbon. For example, the virgin

FIGURE 7-35:

CARBON SLUDGE FILTRATION, LEAF TEST RESULTS
EFFECT OF FORM VACUUM

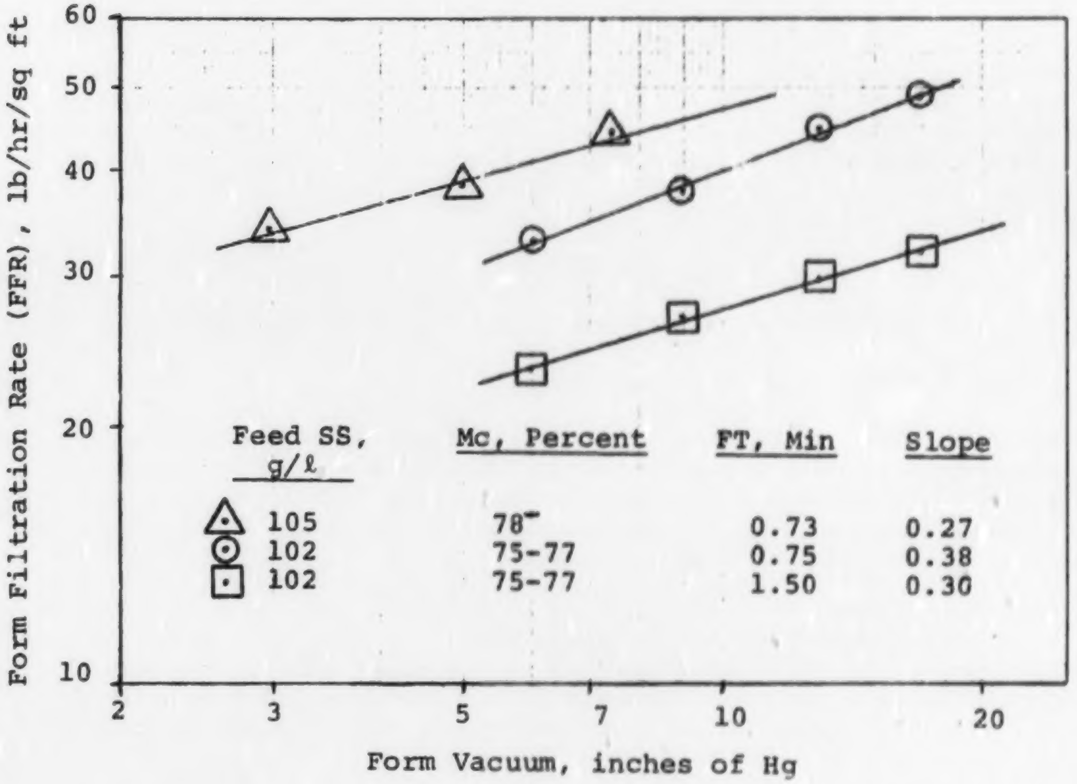
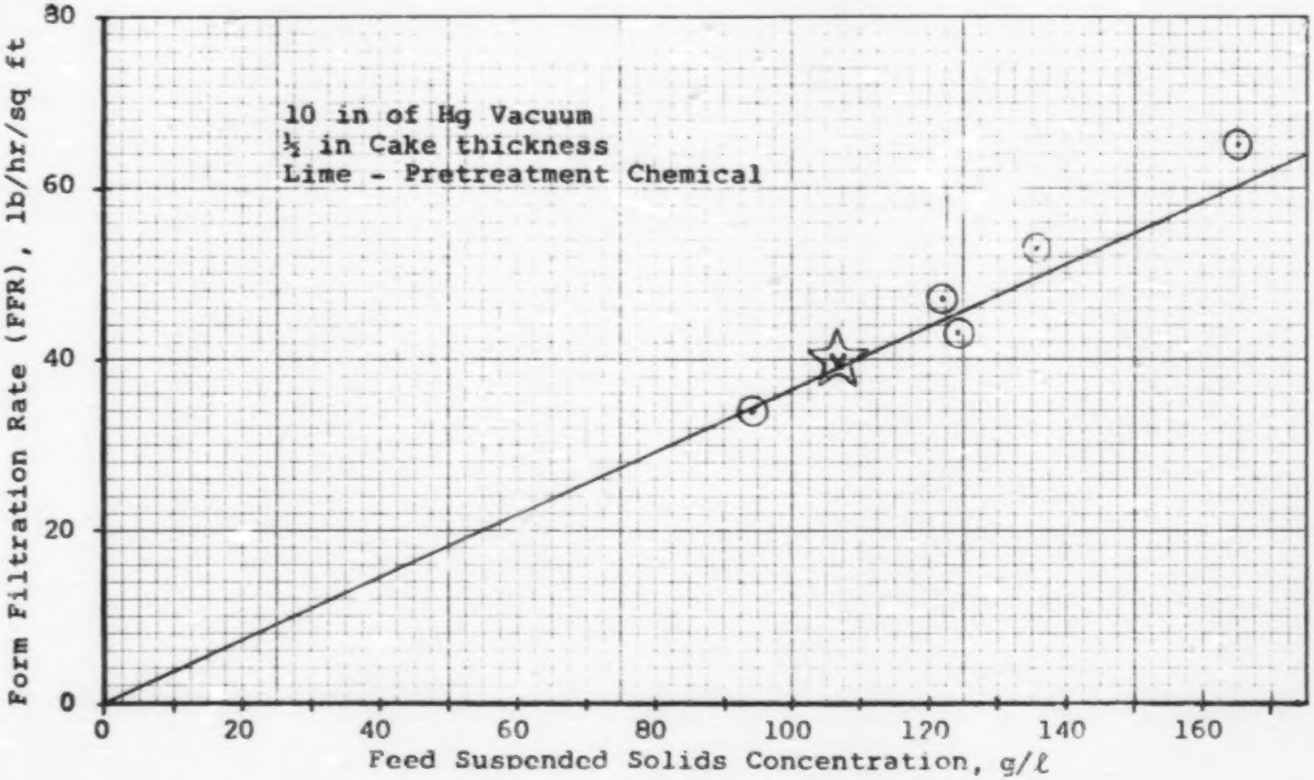


FIGURE 7-36: CARBON SLUDGE FILTRATION, PILOT PLANT RESULTS - EFFECT OF FEED SUSPENDED SOLIDS CONCENTRATION



carbon was 3 percent volatile solids, 3 percent ash and 94 percent fixed carbon; whereas the once spent carbon was 7 percent volatile solids, 7 percent ash and 86 percent fixed carbon.

Inspection of data in Table 7-19 indicates that after Run #40473 a substantial reduction in filtration performance, namely FFR, was experienced. All filter runs after this time were conducted on carbon spent in the pilot plant when alum or ferric chloride were used. During this period of time the carbon was exhausted to higher levels than before. For example, volatile solids in the spent carbon was more than doubled. In addition, lower carbon dosages were being used and higher levels of solids were entering the carbon contactors. The combination of a higher volatile fraction and higher ash (chemical floc) resulted in fixed carbon values of less than 70 percent in the spent carbon.

Results for Runs #50573 and #91473 indicate that a FFR of about 12 lb/hr/sq ft was experienced for a vacuum level of about 11 inches of Hg. Using Equation (1) to predict the effect of vacuum level, a FFR of over 14 lb/hr/sq ft could be expected at a vacuum of 20 inches of Hg. Results of Run #61573, when 6.5 lb/TDS of a cationic polyelectrolyte was used, indicate that a FFR of 15 lb/hr/sq ft was experienced at a vacuum level of 5 inches of Hg.

Extrapolation to 20 inches of Hg would result in predicted FFR of about 23 lb/hr/sq ft. These predictions were based on limited data and thus a substantial factor of safety should be used in any scale-up.

Run #71973 was conducted using cationic polyelectrolyte (Dow C-31) conditioning chemical. The FFR versus FT response for the data obtained during this run was not well correlated. It is apparent that the use of polyelectrolyte at dosages above 6 lb/TDS was effective in increasing FFR.

Run #100873 shown in Table 7-19 was for filtration of regenerated and reused carbon. The FFR is seen to be 14 lb/hr/sq ft for the 3/4 inch cake produced. A single leaf test series on a similar spent (regenerated) carbon sludge indicated a FFR of about 10 lb/hr/sq ft for production of a 1/2 inch thick cake. Both the leaf test and Run #100873 indicated M_c's of 73 percent for θ_D/W values in excess of 1 to 3 minutes.sq ft/lb

Leaf Test Results

Some leaf test results have already been discussed. Table 7-20 presents results from 18 comparisons of pilot plant and leaf test simulations. As found for other sludges, the average pilot plant FFR was significantly higher than the leaf test FFR and the M_c about one percentage point lower.

Summary Discussion

Due to the multiplicity and variance of treatment variables, and in some cases limited data, it would be unwise to generalize about the vacuum dewatering of spent carbon sludge. It can be noted, however, that filter cakes of from 1/4 to 1/2 inch thick were found to be dischargeable. FFR was shown to be linearly related to the feed solids concentration. In addition, FFR was found to be a function of the vaccum level used.

It was also found that FFR and/or M_c were dependent on the type of pretreatment chemical and carbon used in the pilot plant (e.g., virgin or regenerated).

Table 7-21 presents the authors' interpretation of the dewatering results presented and discussed. Prototype operations would most likely be for the case where the most limited data was obtained (i.e.. alum pretreatment and regenerated carbon). Results for these conditions in Table 7-21 indicate that maximization of filter feed solids concentration should be strived for and that the use of a cationic polyelectrolyte should be implemented.

TABLE 7-20: SPENT CARBON SLUDGE FILTRATION: RATIO OF PILOT PLANT TO LEAF TEST RESULTS

Run #	Form Filtration Rate	Filter Cake Moisture Content	Cake Thickness
72872-1	0.97	1.00	1.00
80172-1	0.67	1.01	0.80
80172-2	0.96	0.99	1.00
81672-2	0.97	0.97	1.00
81672-3	0.97	0.97	1.20
81672-4	1.02	1.00	1.00
81672-5	1.21	0.99	1.25
10373-1	1.17	1.01	0.50
10373-2	1.10	1.00	1.50
10373-3	1.13	0.99	4.00
102073-1[a]	0.94	0.99	0.82
102073-2	1.30	0.99	1.00
102073-3	1.38	1.00	1.40
102073-4	1.27	0.96	1.25
102073-5	1.22	0.99	2.00
102073-6	1.62	0.96	1.33
102073-7	1.82	0.99	1.00
102073-8	1.69	1.00	1.33

[a] Virgin Carbon Run

TABLE 7-21: SUGGESTED VACUUM FILTER DESIGN AND PERFORMANCE FOR PAC

Operating Conditions:

Feed SS = 120 g/ℓ
Vacuum = 20 inches of Hg
Effective Submergence = 33 percent
Cake Thickness = 1/2 inch

Pretreatment Chemical	Carbon Source	Yield, lb/hr sq ft	Moisture Content, Percent	θ_D/W min sq ft/ lb
None	Virgin[a]	17	78	≥1-2
Lime	Virgin	18	76-	≥5
Alum or $FeCl_3$	Virgin	7.6	76-	≥5
Alum or $FeCl_3$	Regenerated	5	73	≥3

[a] Not Once Spent

SECTION VIII

POWDERED CARBON REGENERATION

Demonstration of efficient and effective thermal regeneration of carbon using a fluidized bed furnace was a major purpose of this study. There were two main objectives. The first was to demonstrate efficient recovery of fixed carbon. The second was to evaluate the effectiveness of the regeneration process. Effectiveness was evaluated by comparison of carbon characteristics before and after regeneration and by reuse of regenerated carbon in the PAC-PCT pilot plant. It was decided that regeneration effectiveness would not be evaluated until an efficiency of at least 75 percent recovery of fixed carbon was obtained. Because of this priority of objectives, before 75 percent recovery was achieved, the carbon adsorptive characteristics were analyzed for by iodine and molasses numbers only. After 75 percent recovery was achieved, equilibrium adsorption isotherm tests and pilot plant reuse studies were made.

The goal of 75 percent recovery was set because of two factors. First, others had reported that 85 percent, or higher, recovery was possible using a fluidized bed furnace.[2], [3] Secondly, it was shown in the previous contract report that recoveries over 60 percent might make regeneration economically feasible.[1]

During Phase I of this study the priority activity was evaluation of fluidized bed furnace (FBF) operating variables to minimize carbon loss. Because of the unsuccessful results achieved, FBF development activity continued into Phase II. Major changes to the fluidized bed furnace (FBF) established the operating campaigns to be reported here. The first, or Campaign #1, included 3 runs using the furnace as it existed at the end of the previous contract. The bed injection gas (BIG) principal of operation and two different carbon feed points were used. Campaign #2 consisted of 2 runs conducted after the furnace was modified to include an expanding area sand bed chamber, 6 additional gas distribution tuyeres, and an additional higher carbon feed point. Campaign #3 consisted of 6 runs using the furnace as it existed during Campaign #2, but with use of an off-gas recycle (OGR) principal of operation.

Before the furnace could be started, major repair and rework was necessary. During the last run of the previous contract, the burner chamber refractory was melted, thus a new refractory lining was required. This same run resulted in several melted or deformed tuyeres which had to be replaced. A larger burner was installed to allow increased flow capacity. The long shut down between contracts, resulted in considerable corrosion damage to most major electrical instruments and considerable rework was required. A crushed silica sand was obtained locally, and hand-screened to 15 by 30 U.S. Mesh prior to installation. A schematic of the FBF is shown on Figure 8-1.

Results of previous contract studies indicated an unacceptable recovery of carbon. A short period of regenerated carbon reuse had indicated no substantial loss in effectiveness for removing SCOD in the PAC-PCT pilot plant. Toward the end of the previous contract work, the two feed points, shown in Figure 8-1 had been installed, but time was not available for determining their effectiveness. The reason for considering the use of a higher feed point was to allow more time for the bed injection gas to react with excess oxygen entering the sand bed from the firebox.

Studies conducted by others had indicated that lower operating temperatures resulted in lower carbon losses.[2] Therefore, it was decided to operate at as low an operating temperature as allowable with the BIG principal of operation. Because of the need to auto-combust natural gas within the fluidized sand bed, the minimum safe operating temperature was about 1350°F.

CAMPAIGN #1

The first run was conducted using techniques and procedures developed during previous contract work. For this run the 18-inch high feed point was used and the bed temperature was set at 1470°F which was 60°F lower than previous work. The bed fluidized at about 1.2 ft/sec, which was the design capacity of the unit. During the run slight gas and air flow adjustments were made to maintain the stack gas oxygen at less than 0.5 percent by volume.

Table 8-1 presents the FBF operating conditions during the three runs of Campaign #1. The pretreatment chemical refers to the chemical treatment step ahead of the carbon contactors.

FIGURE 8-1: FLUIDIZED BED FURNACE-CAMPAIGN NUMBER 1

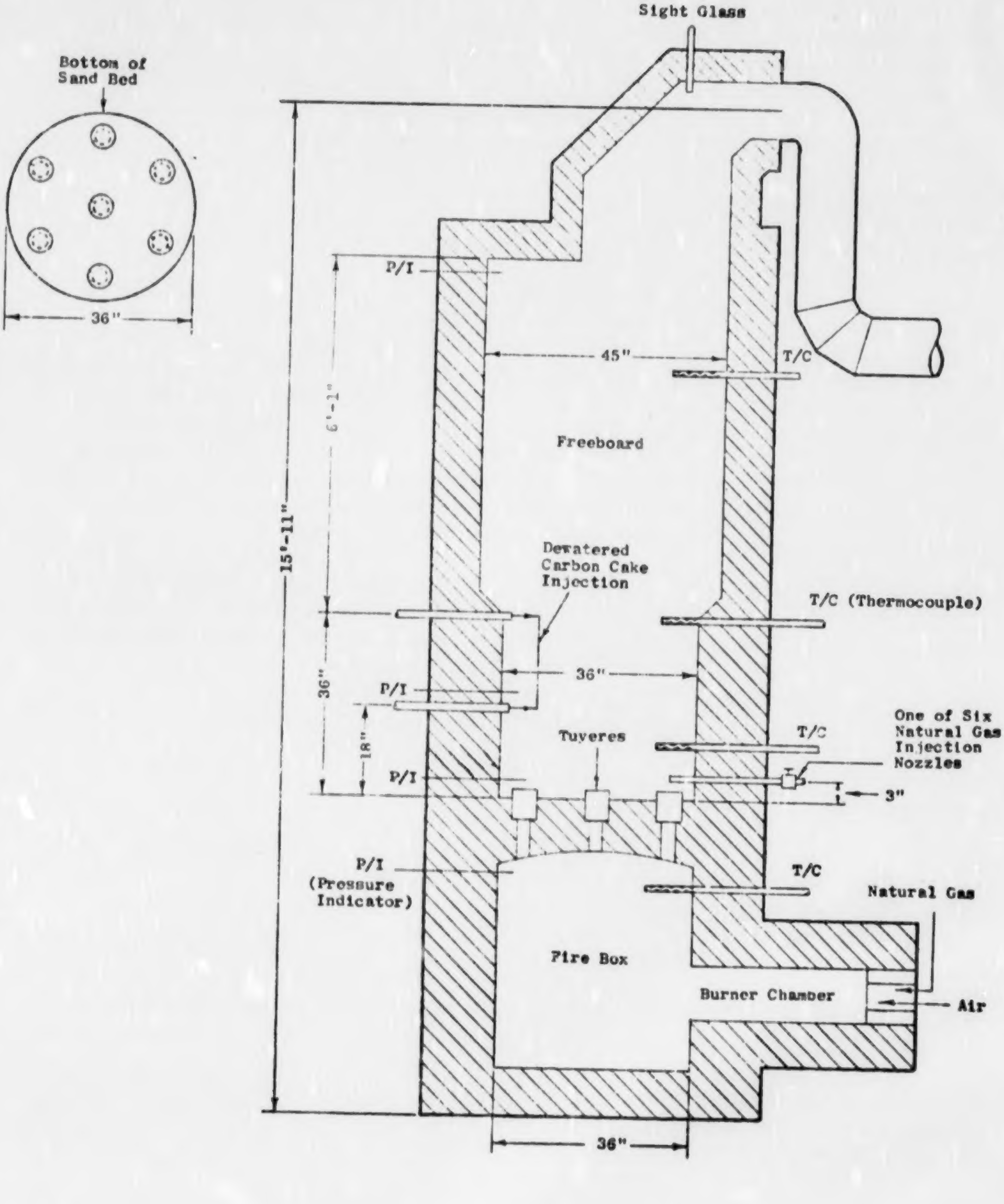

Bottom of Sand Bed
36"
Sight Glass
P/I
45"
T/C
6'-1"
Freeboard
15'-11"
Dewatered Carbon Cake Injection
T/C (Thermocouple)
36"
36"
P/I
18"
Tuyeres
T/C
One of Six Natural Gas Injection Nozzles
P/I
3"
P/I (Pressure Indicator)
T/C
Natural Gas
Fire Box
Burner Chamber
Air
36"

TABLE 8-1: FBF OPERATING CONDITIONS: CAMPAIGN #1

Run Number	1	2	3
Run Date, (1972)	August 16-17,	September 12-13,	September 28-29,
Pretreatment Chemical	Lime	Lime	Lime
Operation Principal	BIG[a]	BIG	BIG
Feed Point (from Bed Floor), in.	18	36	36
Static Sand Bed Depth, in:			
Before run	31	46	40
After run	22	40	36
Sand Size, US Mesh	15x30	15x30	-
Gas Flow, SCFM:			
Burner Air	140	100	120
Burner Gas	3	1.2	1.6
Bed Injection Gas	10	10	>10
Gas Velocity, ft/sec:			
Bottom of Bed	1.2	0.9	1.0
Freeboard[b]	1.2	0.8	0.9
Run Length, hr	22	49	36
Feed Rate, dry lb/hr	65	38	35
Moisture in Feed, % by wgt	77	77	78
Avg Temperature, °F:			
Freeboard	1370	1380	1400
Sand Bed	1470	1470	1490
Firebox	1920	1980	2070
Avg Pressure, in. of H_2O:			
Freeboard	-4	-5	-3
Bottom of Bed	28	38	29
Firebox	42	59	64
Avg O_2 Content of Stack Gas % by volume	0.2	0.2	-

a - bed injection gas
b - includes water vapor

It is identified because 20 to 30 mg/ℓ of suspended solids were present in the overflow from this unit and thus entered the carbon contactors. This concentration of chemical floc is significant when considering that 50 to 300 mg/ℓ carbon was fed to the carbon contactors. Operation principal refers to either bed injection gas, BIG, or off-gas recycle, OGR. Feed point is the elevation above the bottom of the bed that carbon entered the FBF. Sand bed depth was measured directly before start-up and after shut down. During operation, it was measured indirectly by observation of reference points of known height on the inside of the unit. These were viewed through an overhead observation port. This method was used to identify both the initial static and expanded bed depth during operation. Because the freeboard zone was filled with a red glowing carbon haze during carbon feeding, this visual observation could only be made prior to or after carbon feeding. Burner air flow, burner gas flow, and bed injection gas flow were manually adjusted and measured by orifice obstruction manometers. These measurements were sometimes at or beyond the limits of the existing devices. At these times, they were approximated. The values reported in Table 8-1 are arithmetic average of measurements taken about every 1 to 3 hours during a run. Gas velocities were computed and adjusted to operating temperature. The gas velocity in the freeboard includes water vapor contributed by the feed. This was significant at times. Run length is actual feed time or clock time, minus any down time.

Percent moisture in the feed was obtained from solids analysis of vacuum filter cake samples. Temperatures were measured within the bed by thermocouples and continuously printed on a multi-point recorder. Pressures were measured by diaphragm type pressure meters.

Table 8-2 presents the carbon recovery and characteristics results of the three runs during Campaign #1. Fixed carbon recovery during Run #1 was only 49 percent. A large quantity of ash appeared in the product, 35 percent by weight. Almost half of this ash was identified as being sand. Sand loss from the furnace during both prerun start-up and during the run amounted to about 30 percent by volume. Some of this sand loss could be accounted for by intermittent high velocities attained prior to the run, the high attrition rate of the locally obtained sand, and the large, almost explosive, volume expansion caused by the water in the carbon feed. The volume of flow attributed to the water vapor was half of that of fluidizing gas streams in the freeboard. Since the carbon was fed at a single point, rapid vaporization of this large amount of water may have

TABLE 8-2: CARBON CHARACTERISTICS[a]: CAMPAIGN #1

Run Number	1		2		3	
Run Date	August 16-17, 1972		September 12-13, 1972		September 28-29, 1972	
Fixed Carbon Recovery %	49		62		56	
	Spent	Regenerated	Spent	Regenerated	Spent	Regenerated
Solids, % by wgt						
Volatiles	5.1	1.4	12.8	1.2	3.6	0.7
Ash	9.9	34.6	6.4	23.0	5.0	15.8
Fixed Carbon	85	64.1	80.8	75.8	91.4	83.5
Iodine Number	776	832	700	862	796	1020
Molasses Number	16	41	18	44	44	201
Acid Solubles, % by wgt	--	20.1	--	--	--	--

[a] See Table 8-8 for virgin carbon properties

had a significant effect on the travel of the carbon; probably causing short circuiting of carbon to the surface of the bed. If this phenomenon existed it could have resulted in sand being literally thrown out of the unit.

The computed feed rate of 65 lb/hr during Run #1 was consistent with previous experience.

To give even more time for the oxygen in the gas flow from the firebox to be consumed by the bed injection gas, it was decided to move the feed point up another 18 inches, to the 36-inch feed point during Run #2. This seemed to be the only variable that could be changed to increase the recovery from that of Run #1. Sand was added to the unit to a total depth of 46 inches. This insured that the feed point would be covered during operation, even if some sand were lost during start-up. Because of the high sand losses during Run #1, it was decided to drop the fluidization velocity during Run #2 to 0.9 ft/sec.

The recovery of fixed carbon during Run #2 was better than for Run #1, but still only 62 percent, as shown in Table 8-2. The sand losses were lower, 19 percent by volume, with a corresponding reduction of ash in the product, 23 percent by weight. Freeboard velocities of 1.0 ft/sec or less, if reasonably uniform, should not have carried sand out of the unit. The carbon feed rate dropped to 38 lb/hr during Run #2.

A substantial spent carbon inventory within the pilot plant units necessitated that the third run be made. There was no rationale for changing operational variables, therefore, it was decided to conduct Run #3 using the same conditions as for Run #2. A decision had been made just prior to Run #3 to seek expert advice on the FBF design and operation in hopes of establishing some meaningful direction to pursue further obviously needed development work.

Recovery of fixed carbon during Run #3 was 55 percent. Sand loss was low, 10 percent by volume, with a corresponding smaller ash content in the product, 16 percent by weight. This lower level of sand loss was possibly due to having purged out fines during the previous runs.

It was decided at this point that because sand seemed to be continuously carried over into the product carbon even at low fluidization velocities, a simple hydrocyclone should be installed on the scrubber underflow. This was done and during subsequent runs it performed very well.

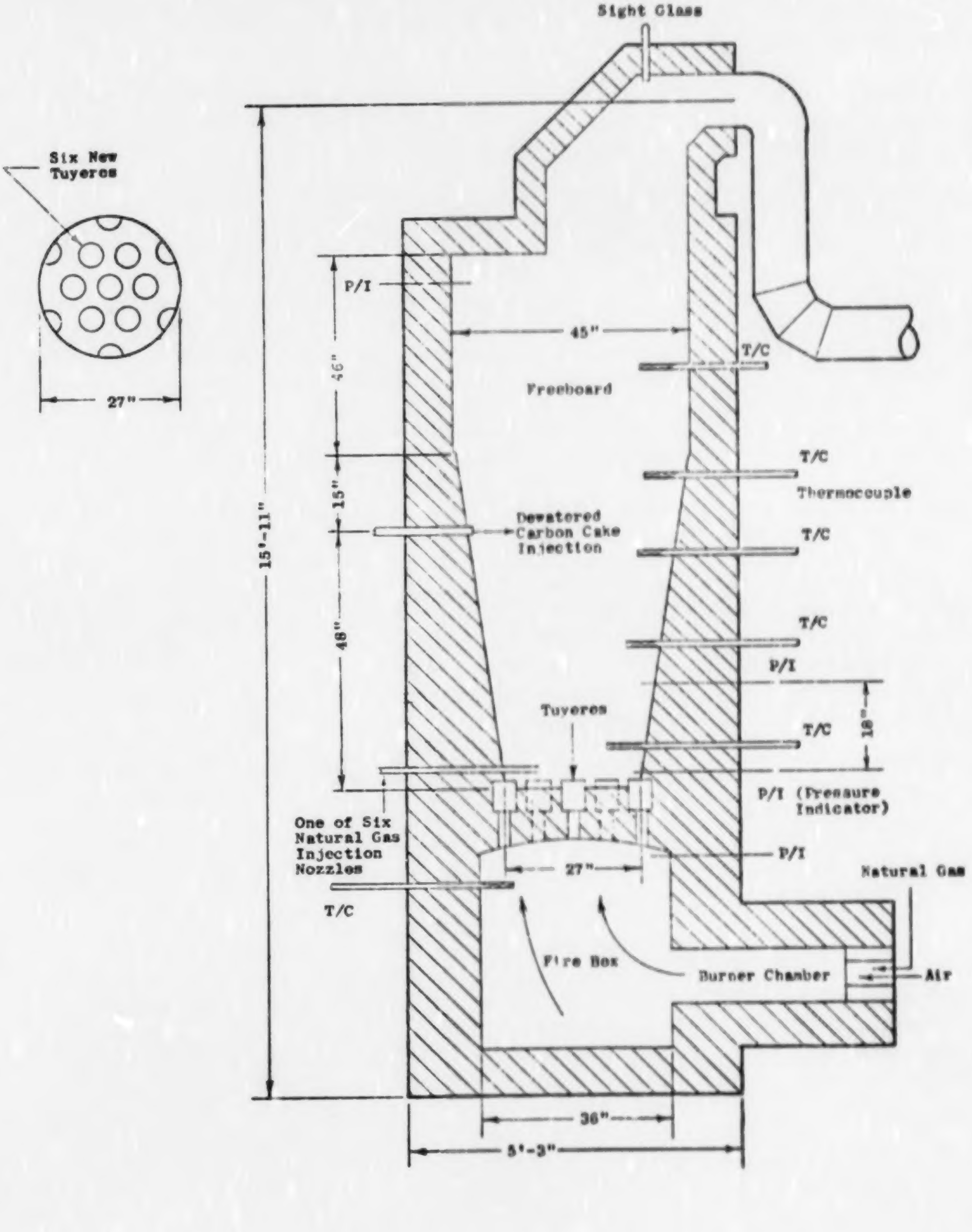
During the three runs of Campaign #1, freeboard temperatures were 90 to 110°F below those for the sand bed. Thus, it was unlikely that burning of substantial amounts of either carbon or natural gas occurred above the bed. This would infer that carbon was burning within the bed, and that an inadequate amount of oxygen was being consumed prior to the carbon feed point.

Consultation with others indicated that inadequate mixing was taking place in the FBF sand bed and that mixing could be improved by increasing fluidization velocity. Since the burner equipment was sized for the existing furnace, with an approximate maximum velocity of 1.2 ft/sec at the bottom of the bed, and since high sand losses were being experienced at these velocities it was decided that the bed should be reshaped to that of an inverted truncated cone. This shape would provide a small area at the bottom of the bed with a resultant higher fluidization velocity and increased mixing at a given volumetric flow rate. The area at the freeboard would be the same as before and would thus have the same tendency for sand loss. The decrease in the bottom area allowed for an increase in velocity of approximately 80 percent. The area change was from 7.08 sq ft to 3.95 sq ft. To get better ditribution of the fluidizing gas it was recommended by others that six additional tuyeres be installed in the bed floor. This new tuyere pattern is shown on the inset of Figure 8-2. The outside 6 tuyeres contained 3 orifices each and the remaining 7 tuyeres contained 6 orifices each. These orifices were 0.272 inches in diameter. This allowed for good sand retention with no gas flow and a small headloss at operating gas flows. An operating headloss of 10 inches of water was originally designed for.

Additional changes made to the furnace at this time were a new feed point at the 4-ft bed depth level, 2 new thermocouples (one at the 4-ft level and one at the 5-ft level) and a larger sight glass on the top of the furnace. The final furnace configuration prior to Campaign #2 is shown on Figure 8-2.

It was also recommended by others that a high silica sand with a wide particle size distribution be used in place of the uniform (15 by 30 U.S. Mesh) locally obtained sand. Thus about 3-ft of recommended 16 by 80 U.S. Mesh high silica sand was procured and installed prior to Run #4.

FIGURE 8-2: FLUIDIZED BED FURNACE-CAMPAIGN NUMBER 2


Six New Tuyeres
27"
Sight Glass
15'-11"
P/I
46"
45"
T/C
Freeboard
15"
T/C
Thermocouple
Dewatered Carbon Cake Injection
T/C
48"
T/C
P/I
Tuyeres
T/C
18"
P/I (Pressure Indicator)
One of Six Natural Gas Injection Nozzles
27"
P/I
Natural Gas
T/C
Fire Box
Burner Chamber
Air
36"
5'-3"

CAMPAIGN #2

After a delay of several months due to late deliveries of materials and to other project priorities, two runs were conducted using the BIG principal of operation and the modified FBF. The new higher carbon feed point (at 48 inches) was used to allow maximum time for the excess oxygen to be consumed. Combustion air and gas were adjusted to provide a recommended 2.5 fps sand bed bottom velocity. The furnace operating conditions and results for Runs #4 and #5 are presented in Tables 8-3 and 8-4, respectively. The fixed carbon recovery results in Table 8-4 indicate the very poor recoveries experienced.

Prior to Run #4 about 34 inches of 16 by 80 U.S. Mesh high silica sand was placed in the FBF. Visual observation of the fluidized bed prior to and after carbon feeding during this run indicated that the bed was fluidized to about the 4-ft level. After Run #4 only about 24 inches of sand was left in the FBF. Most of the sand blown out of the FBF was caught in the newly installed hydrocyclone separator. The substantial increase in ash indicated in Table 8-4 suggests that copious amounts of fine sand passed through the hydrocyclone separator.

Prior to Run #5 two additional feet of 16 by 80 U.S. Mesh high silica sand was added to the FBF. At the lower fluidization velocity indicated in Table 8-3, considerably less sand was lost. Of the 19 percent ash found in the regeneration product, some 8.4 percent was found to be acid soluble. Thus, a nominal amount of sand found its way into the regenerated carbon recovery system. This observation was substantiated when upon passing the regenerated carbon through a 60 U.S. Mesh screen prior to reuse in the pilot plant, very little sand was collected.

Average freeboard and sand bed temperature in Table 8-3 indicate that during Run #5 substantial heat was generated in the freeboard region. Presumably either carbon or natural gas and oxygen was present above the fluidized bed. This observation was not found for Run #4. The higher fluidization velocity and/or the shallower bed during Run #4 may have been the reason.

Inspection of Table 8-4 shows that almost full recovery of iodine and molasses numbers were achieved during Runs #4 and #5. Figure 8-3 shows the regenerated carbon that was recovered had SCOD removal properties essentially equal to virgin carbon.

TABLE 8-3: FBF OPERATING CONDITIONS: CAMPAIGN #2

Run Number	4	5
Run Date, (1973)	April 4-5,	May 2-3,
Pretreatment Chemical	Lime	Alum
Operation Principal	BIG[a]	BIG
Feed Point (from Bed Floor), in.	48	48
Static Sand Bed Depth, in:		
Before run	34	48
After run	24	42
Fluidized Bed Depth, in.	48-60	-
Sand size, US Mesh	16x80	16x80
Gas Flow, SCFM:		
Burner Air	140	120
Burner Gas	2.5	-
Bed Injection Gas	>10	8.7
Gas Velocity, ft/sec:		
Bottom of Bed	2.5	1.9
Freeboard[b]	1.3	∿1.0
Run Length, hr	12.5	19.2
Feed Rate, dry lb/hr	71	58
Moisture in Feed, % by wgt	75	74
Avg Temperature, °F:		
Freeboard	1190	1480 ± 100
Sand Bed	1390	1380 ± 100
Firebox	1960	1900
Avg Pressure, in. of H_2O:		
Freeboard	-3	-4
Bottom of Bed	36	-
Firebox	44	53
Avg O_2 Content of Stack Gas, % by volume	0.2	0.1

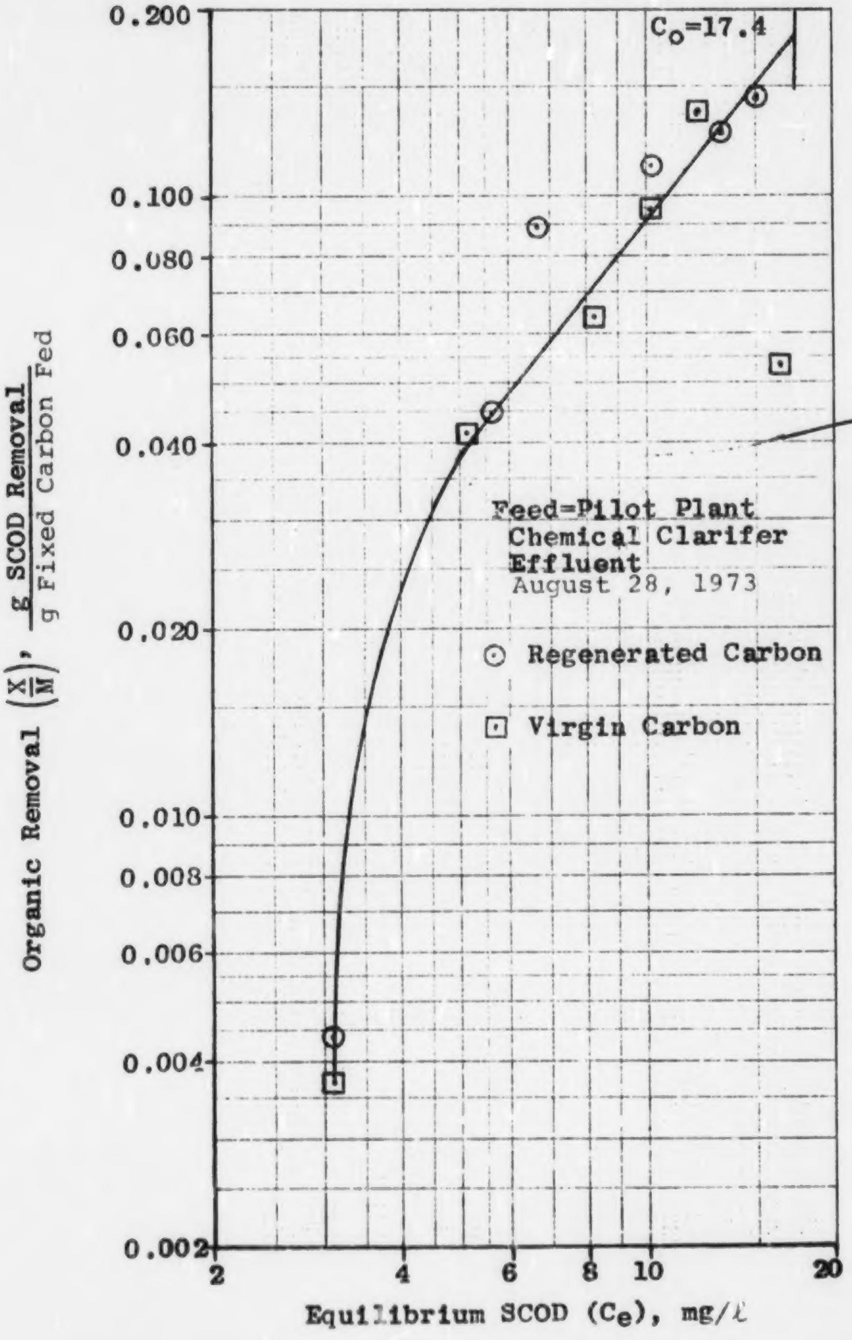
a - bed injection gas
b - includes water vapor

TABLE 8-4: CARBON CHARACTERISTICS[a]: CAMPAIGN #2

Run Number	4		5	
Run Date	April 4-5, 1972		May 2-3, 1972	
Fixed Carbon Recovery %	39		36	
	Spent	Regenerated	Spent	Regenerated
Solids, % by wgt				
Volatiles	5.7	0.3	6.6	1.2
Ash	6.9	44.7	7.3	19
Fixed Carbon	87.4	55.1	86.1	79.8
Iodine Number	679	988	588	955
Molasses Number	25	108	5	77
Acid Solubles, % by wgt	--	--	--	8.4
Aluminum, % by wgt	--	--	--	1.6

[a] See Table 8-8 for virgin carbon properties

FIGURE 8-3: COMPARATIVE EQUILIBRIUM ADSORPTION ISOTHERM TESTS: REGENERATION RUN NUMBER 5



MODIFICATION OF FBF OPERATING PRINCIPAL

After having exhausted all reasonable alterations of physical and operation variables, including those recommended by others, the contractor concluded that the BIG principal of FBF operation would not work for the unit being used. This conclusion was officially stated after Run #4. At this time (late April, 1973) only about three months of pilot plant operating funds were left in the contract. Reuse of regenerated carbon had not yet been attempted!

Based on successful results achieved by others using a similar FBF regeneration system (reference 2), but with a fundamentally different method of oxygen control, the contractor decided to modify the existing unit in an attempt to, at least, demonstrate that high carbon recovery was possible. It was recognized that little if any regenerated carbon reuse could be achieved with the limited time available. Because of this fact negotiations were undertaken with the EPA to explore the possibility of an extension to the contract. A 3 month extension was obtained with the major purpose being to demonstrate high carbon recovery and to evaluate the effectiveness of regeneration by reuse of carbon in the pilot plant.

The relatively minor modifications in the FBF unit required to allow operation using the off-gas recycle, OGR, principal can be seen by comparing Figure 8-4 with 8-2. Sections IV and V contain additional information about the modification and the approach to operation with the OGR principal.

CAMPAIGN #3

This Campaign spanned a 4-1/2 month period from June to November, 1973, which included the 3 month contract extension. During this period six regeneration runs were made, one using virgin carbon feed. Tables 8-5 and 8-6 present FBF operating conditions and carbon characteristics for the five runs using spent carbon. The only major operational variations were sand bed temperature and fluidization velocity. Operating conditions were set immediately prior to any given run and were based on feedback from previous runs. The following is a brief discussion of each run, indicating the rationale used to set operating conditions, fixed carbon recovery, carbon characteristics and any anomalies or problems encountered.

Run #6 was the first using the OGR operating principal. Relatively low fluidization velocity and sand bed temperature were used in hopes of maximizing fixed carbon recovery.

FIGURE 8-4: FLUIDIZED BED FURNACE-CAMPAIGN NUMBER 3

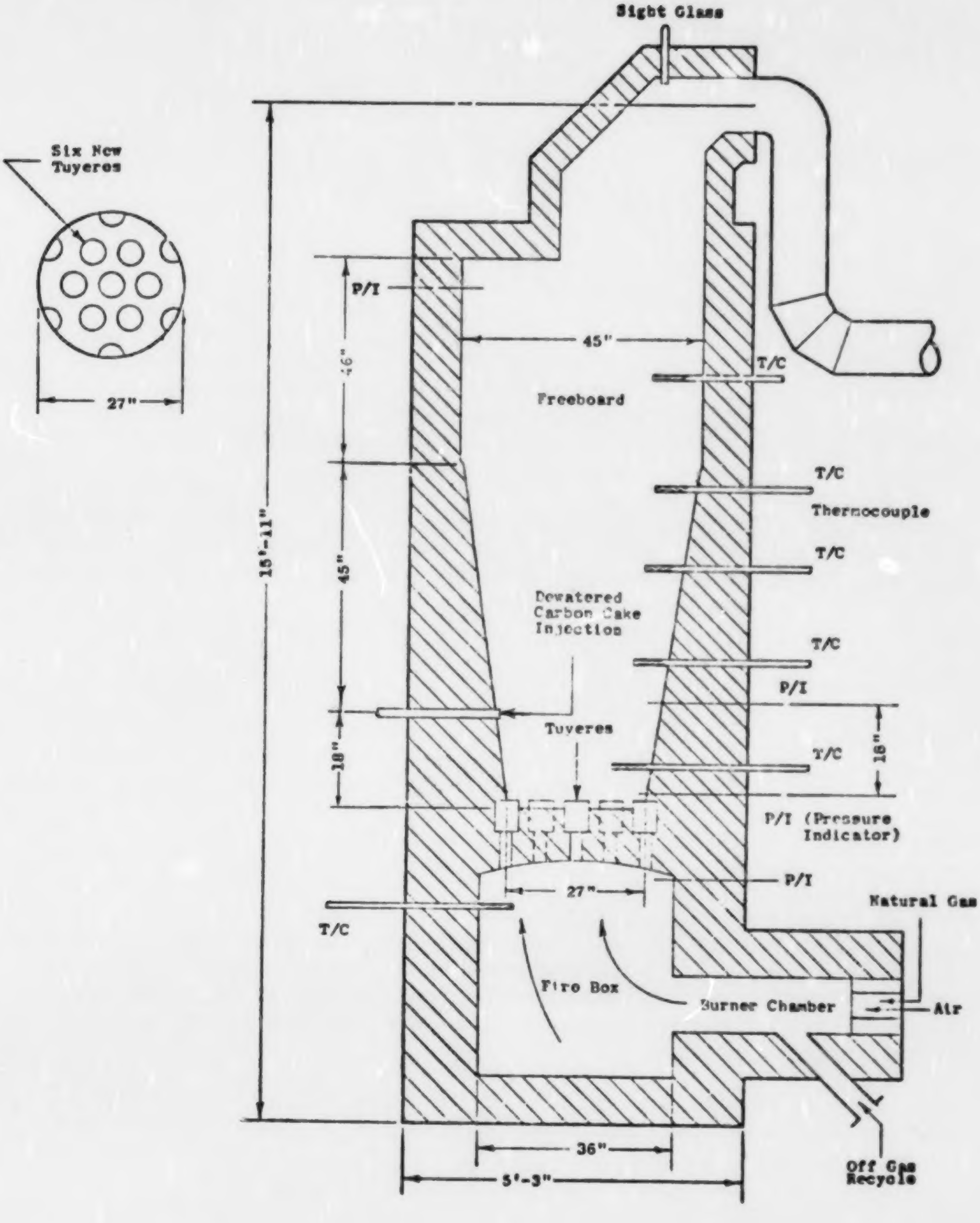

Sight Glass
Six New Tuyeres
27"
15'-11"
P/I
46"
45"
45"
18"
T/C
Freeboard
T/C
Thermocouple
T/C
Dewatered Carbon Cake Injection
T/C
P/I
Tuyeres
T/C
18"
P/I (Pressure Indicator)
27"
P/I
T/C
Natural Gas
Fire Box
Burner Chamber
Air
36"
5'-3"
Off Gas Recycle

TABLE 8-5: FBF OPERATING CONDITIONS: CAMPAIGN #3

Run Number	6	7	8	9	11
Run Date, (1973)	June 17-18,	July 18-20,	Sept 11-13,	Oct 8-10,	Nov 2-3,
Pretreatment Chemical	Alum	$FeCl_3$	$FeCl_3$	$FeCl_3$	Alum
Operation Principal	OGR[a]	OGR	OGR	OGR	OGR
Feed Point (from Bed Floor), in.	18	18	18	18	18
Sand Bed Depth, in:					
Static	32	32	36	-	-
Fluidized	-	44+	48+	-	-
Sand Size, US Mesh	16x30	16x30	16x30	16x30	16x30
Gas Flow, SCFM:					
Burner Air	80	90	85	90	93
Burner Gas	8	9	9.2	10	10
Off Gas Recycle	70	60	75	80	90
Gas Velocity, ft/sec:					
Bottom of Bed	2.0	2.2	2.2	2.7	2.9
Freeboard[b]	0.8	1.0	1.0[c]	1.1	1.1
Run Length, hr	30.7	28.3	21.5	51.7	46.0
Feed Rate, dry lb/hr	21	27	37	16	13
Moisture in Feed, % by wgt	76	76	76[a]	73	72
Avg Temperature, °F:					
Freeboard	1130	1330	1170	1470	1470
Sand Bed	1250	1400	1250	1550	1550
Firebox	1900	1970	1960	2000	2000
Avg Pressure, in. of H_2O:					
Freeboard	-4	-3	-3	-1	0
Firebox	57	64	72	70	68
Avg O_2 Content of Stack Gas, % by volume	0.2	0.3	0.3	0.0	0.1

a - off gas recycle
b - includes water vapor
c - estimated

TABLE 8-6: CARBON CHARACTERISTICS[a]: CAMPAIGN #3

Run Number	6		7		8		9		11	
Run Date	June 17-18, 1973		July 18-20, 1973		September 11-13, 1973		October 8-10, 1973		November 2-3, 1973	
Fixed Carbon Recovery %	100		78		93		100		76	
	Spent	Regenerated	Spent	Regenerated	Spent	Regenerated	Spent	Regenerated	Spent	Regenerated
Solids, % by wgt										
Volatiles	8.7	1.5	12.7	3.0	16.0	4.5	13.4	1.8	13.1	2.8
Ash	9.2	16.5	15.5	17.7	19.1	26.0	18.0	32.7	23.3	26.0
Fixed Carbon	82.2	82.0	71.8	79.4	69.9	69.5	68.6	65.5	63.6	71.2
Iodine Number	506	788	311	684	158	527	266	723	274	662
Molasses Number	15	59	10	69	7	28	6	56	8	58
Acid Solubles % by wgt	--	9.7	--	10.2	--	19.0	--	16.1	--	15.8
Iron,% by wgt	--	--	--	3.2	--	12.4	--	7.0	--	5.7
Aluminum, % by wgt	--	--	--	--	--	--	--	0.4	--	18.8

[a]See Table 8-8 for virgin carbon properties

Results in Table 8-6 indicate that 100 percent recovery of fixed carbon was achieved.

Operation of the FBF was very stable during Run #6. The increase in ash which was not acid-soluble ash in the regeneration product compared to the spent carbon (Table 8-6) indicated that some sand was probably still being lost from the FBF unit and not being captured by the hydrocyclone.

Comparison of regenerated carbon iodine and molasses numbers with previous runs indicated less than full recovery of these properties. A comparative equilibrium adsorption isotherm with virgin carbon (shown on Figure 8-5) also indicated a significant loss of SCOD adsorptive capacity. It was assumed that the low sand bed temperature of 1250°F was the cause of not recovering full adsorptive capacity. Thus, it was decided to increase the sand bed temperature during the next Run, #7.

Table 8-6 indicates that only 78 percent recovery of fixed carbon was achieved during Run #7. Comparison of Run #7 and #6 operating conditions indicate only two significant differences. First, the sand bed temperature was increased from 1250 to 1400°F. Secondly, the carbon contacting system feed pretreatment chemical was different, being alum for Run #6 and $FeCl_3$ for Run #7. It was not considered likely that a 150°F increase in sand bed temperature would have caused such a drastic reduction in fixed carbon recovery. Thus, it was conjectured that the presence of ferric floc and/or adsorbed iron (see Section VI, Operating Period IV) might have catalyzed the burning of fixed carbon. To eliminate the possible temperature effect the next Run, #8, would be run at the lower temperature of 1250°F.

Analysis of spent carbon characteristics in Table 8-6 and comparison with previous regeneration runs, indicated that carbon was finally being exhausted to a high degree. Volatile solids, iodine and molasses numbers all indicated a relatively high degree of carbon exhaustion. Regenerated carbon characteristics showed a substantial recovery of adsorptive properties, but not full recovery. A comparative equilibrium adsorption isotherm with virgin carbon (see Figure 8-6) showed a significant loss of SCOD adsorptive capacity at equilibrium SCOD values of 10 to 15 mg/ℓ.

During Run #7 little sand was lost from the FBF and only a nominal amount appeared in the recovered product.

FIGURE 8-5: COMPARATIVE EQUILIBRIUM ADSORPTION ISOTHERM TESTS: REGENERATION RUN NUMBER 6

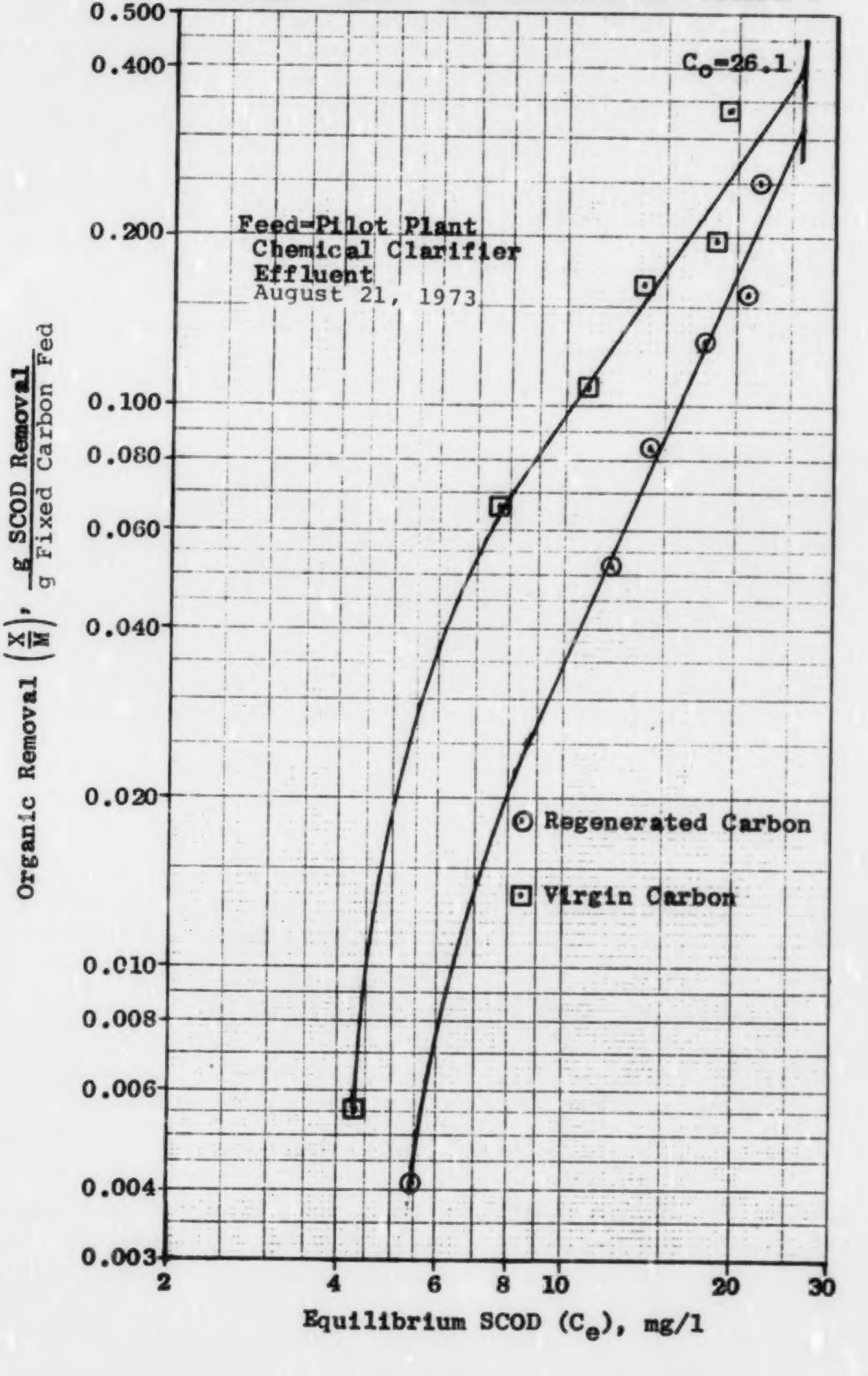
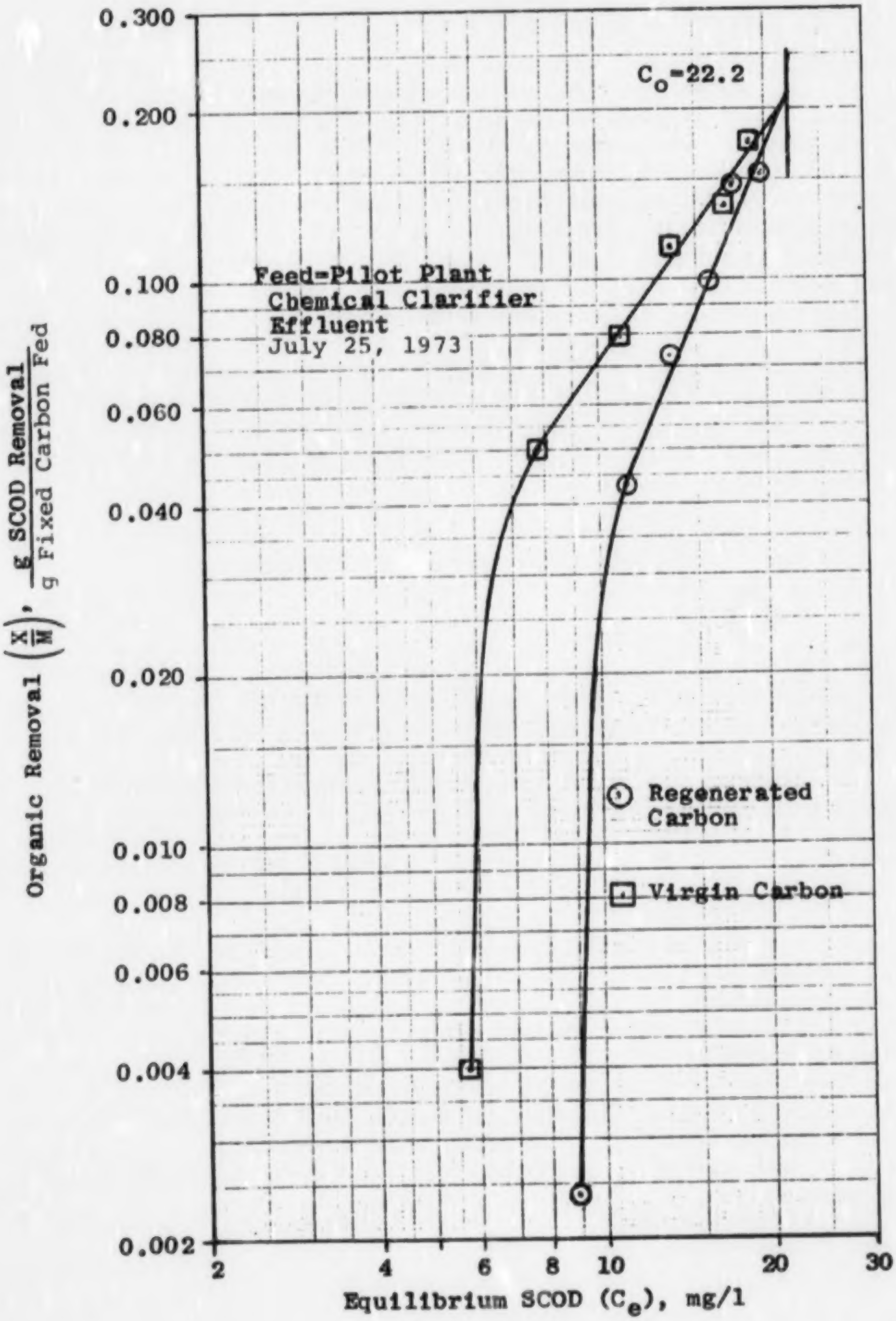


FIGURE 8-6: COMPARATIVE EQUILIBRIUM ADSORPTION ISOTHERM TESTS: REGENERATION RUN NUMBER 7



Though the regenerated carbon adsorptive properties were less than 100 percent recovered, it was still decided to reuse the product in the PAC-PCT pilot plant. Thus, in mid-August regenerated carbon from Runs #5, #6 and #7 were mixed together and used for carbon treatment system feed.

The start of Run #8 was planned for August 16. However, on August 15, with the FBF at operating temperature, an abnormally high pressure drop of 150 inches of water developed across the tuyeres and sand bed. The combustion air blower flow was reduced to only about 50 SCFM. It was feared that either the tuyeres had sustained major damage or that the sand bed had been "cemented". The latter suspicion stemmed from the observation that after Run #7, which was the first using a ferric chloride pretreatment spent carbon, the sand bed grains were uniformly coated with a dark reddish-brown precipitate. This coating could have been either iron oxide or iron phosphate. It was felt that had momentary general or localized sand temperature exceeded about 1600°F, that iron phosphate fusion might have occurred and resulted in cementation of sand grains around or in the orifices of the tuyeres.

Because of the above mentioned high pressure drop problem, the FBF had to be cooled down to allow removal of sand and inspection of the tuyeres. No obvious signs of tuyere damage, orifice clogging or sand grain fusion were found! The sand bed was put back into the FBF, with about 4 inches of new 16 by 30 U.S. Mesh high silica sand, and the unit started-up again. Pressure drops returned to normal, however, full combustion air flow was still not obtained!

The air blower was dismantled and considerable packing material between the multiple stages found missing. The blower was repaired, the FBF brought back up to operating temperature and Run #8 started - after about a 4 week loss of time.

Recall that Run #8 was to be conducted at the reduced bed temperature of 1250°F to observe if the poor fixed carbon recovery during Run #7 might be due to the higher bed temperature (1400°F) or possibly due to the presence of substantial amounts of iron in the spent carbon. As seen in Table 8-6, a good recovery of 93 percent fixed carbon was achieved. It was, therefore, concluded that the presence of iron compounds in the spent carbon did not drastically effect fixed carbon recovery, at least at a 1250°F operating temperature.

Iodine and molasses numbers for Run #8 indicated that the spent carbon was exhausted to a high degree. They also indicated that, though a substantial recovery was experienced, the regenerated carbon values were very low. A comparative equilibrium adsorption isotherm test again indicated substantial loss of SCOD adsorptive capacity (see Figure 8-7).

Since regenerated carbon was being reused in the pilot plant, and poor SCOD adsorptive properties could be detrimental to performance, it was decided to shift priorities from seeking to maximize fixed carbon recovery to maximizing recovery of adsorptive properties. Thus, a higher sand bed temperature was scheduled for Run #9.

In addition to a higher operating temperature, the fluidization velocity was increased. Table 8-6 shows that a fixed carbon recovery of 100 percent was achieved. Recovery of adsorptive properties was judged to be very good, considering that the spent carbon feed was once regenerated and reused carbon. Figure 8-8 shows SCOD removal results.

The high recovery of fixed carbon for Run #9 seemed to negate previously [illegible]ssed concern about the presence of iron compounds in [illegible] carbon and that higher operating temperatures might be causing increased carbon losses. Apparently other operational or process variations were, at least in part, the cause of inconsistent recoveries of fixed carbon and possibly adsorptive properties.

Run #11 was conducted during the final week of pilot plant operation. The spent carbon feed was a mixture of once and twice regenerated and reused carbon with alum pretreatment chemical. Table 8-6 shows that only 76 percent recovery of fixed carbon was achieved. Some carbon was burnt during the run because of a malfunctioning main gas line control valve. On at least two occasions, this valve closed and the combustion air blower remained on. This resulted in air, with high oxygen content, going directly to the sand bed. Carbon feeding continued at a decreasing rate for several minutes until the bed temperature reached a level which shut the feed pump off. The amount of carbon burned due to this problem was undeterminable, but probably less than 2 to 4 percent of the total fixed carbon fed during the run.

Observation of carbon characteristics in Table 8-6 and the equilibrium adsorption isotherm test results on Figure 8-9 indicates that substantial recovery of adsorptive properties was achieved. Especially when considering that a mixture

FIGURE 8-7: COMPARATIVE EQUILIBRIUM ADSORPTION ISOTHERM TESTS: REGENERATION RUN NUMBER 8

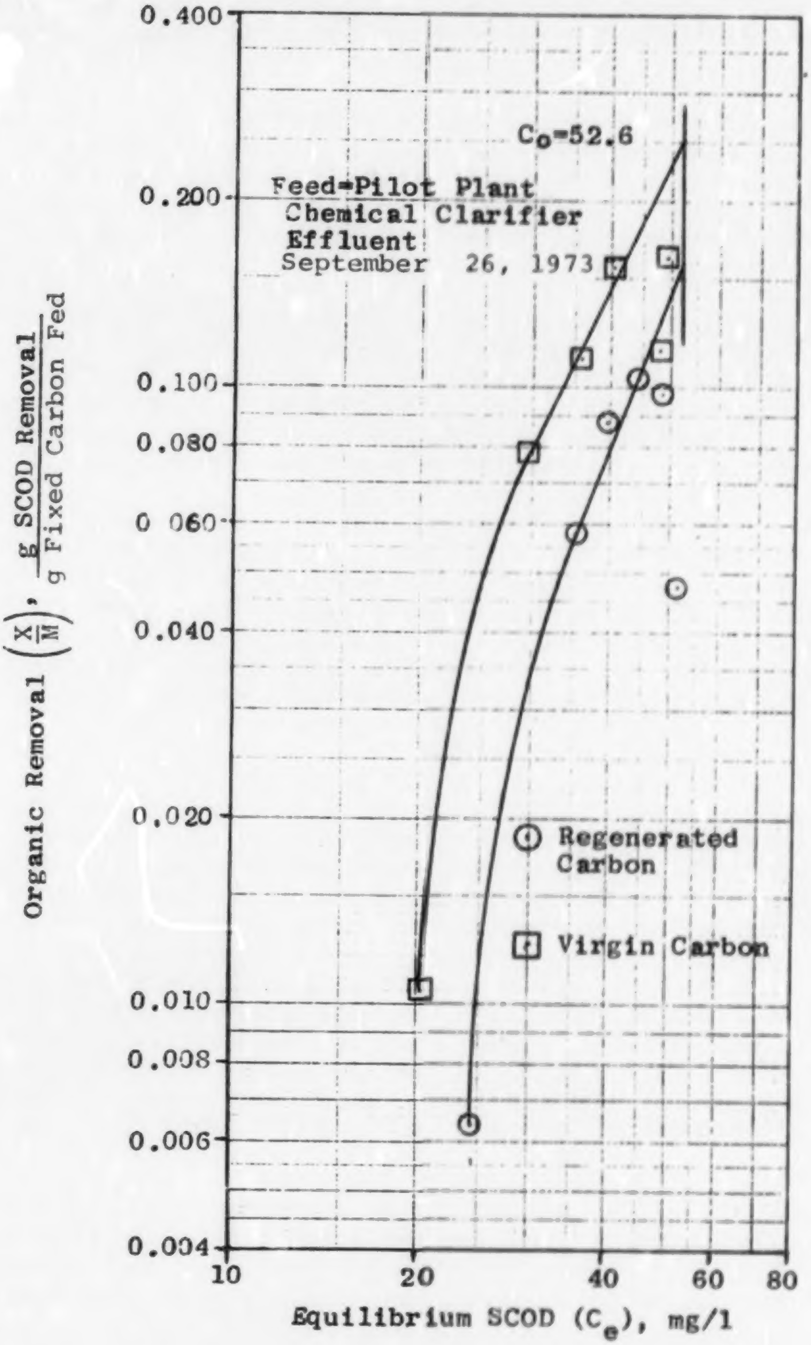


FIGURE 8-8: COMPARATIVE EQUILIBRIUM ADSORPTION ISOTHERM TESTS: REGENERATION RUN NUMBER 9

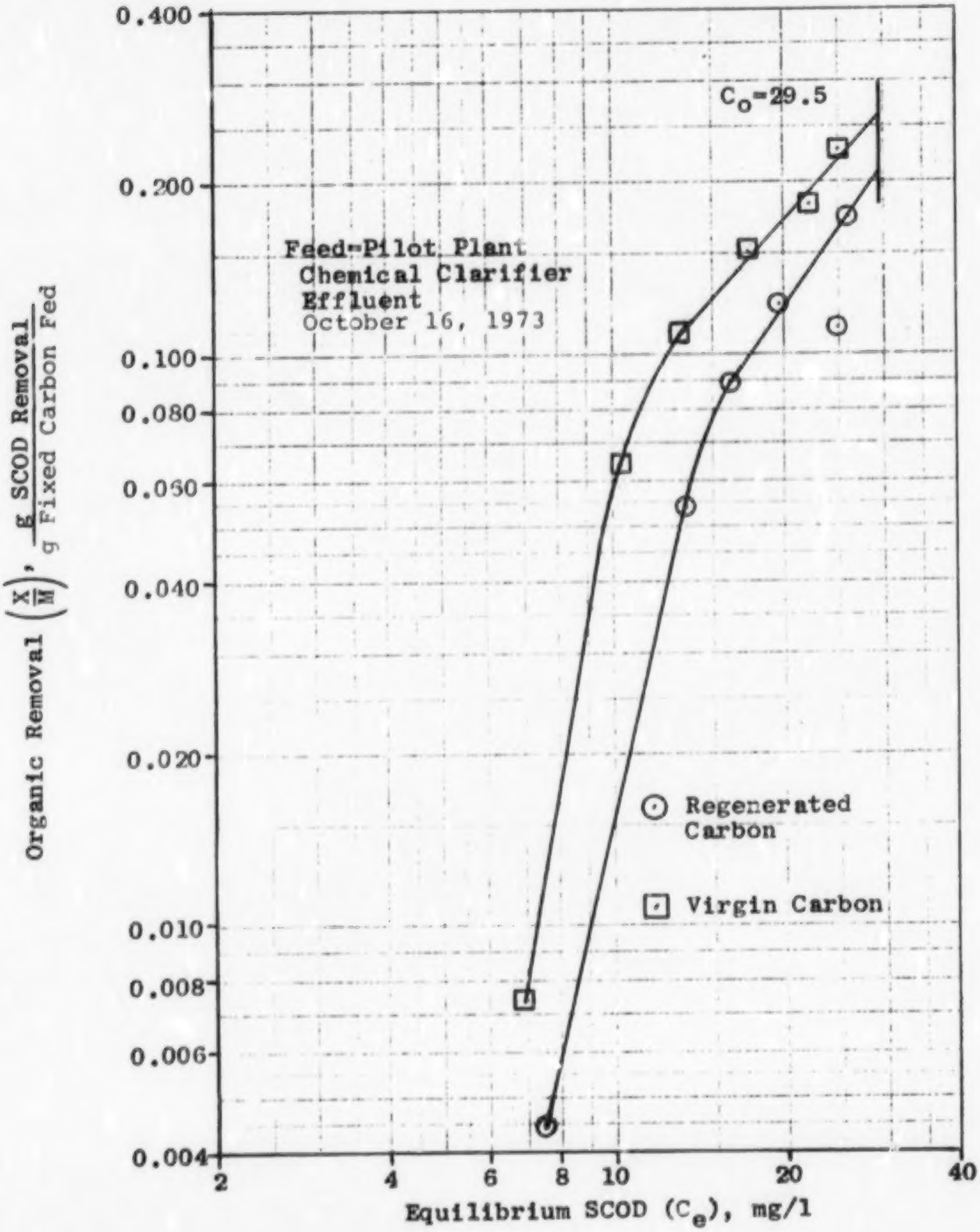
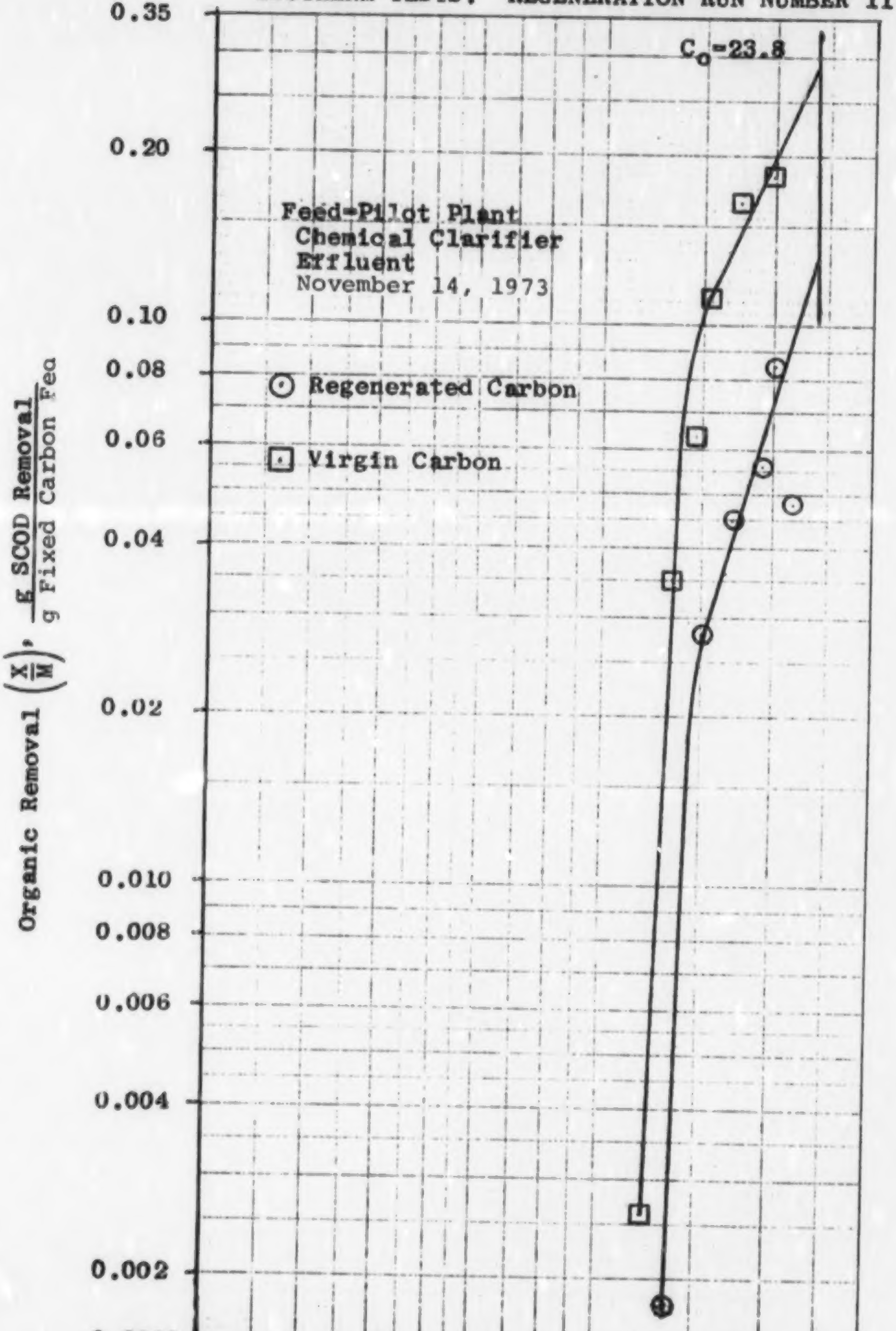


FIGURE 8-9: COMPARATIVE EQUILIBRIUM ADSORPTION ISOTHERM TESTS: REGENERATION RUN NUMBER 11



of once and twice regenerated carbon had been spent in the pilot plant carbon contacting system.

A special regeneration run was conducted October 20, 1973, using vacuum filtered virgin carbon as the regeneration system feed. The FBF operating conditions and results for this Run #10, are presented in Tables 8-7 and 8-8. FBF operating conditions are seen to be similar to those for Runs #9 and #11 except that the sand bed temperature was 100°F lower, 1450°F. Table 8-8 shows excellent fixed carbon recovery of 98 percent. No significant change in adsorptive properties were found for molasses number and SCOD adsorption tests, Figure 8-10. It is interesting to note that an increase in iodine number from 942 to 1040 was experienced. It is normally expected that thermal regeneration of carbon results in a reduced iodine number due to collapsing or blockage of micropores.

DISCUSSION OF RESULTS

Evaluation of the results of this study indicate wide variations in the effectiveness of the regeneration system. Variability of the degree of carbon exhaustion, pretreatment chemical used, FBF operating conditions (fluidization velocities, temperature, sand bed depth, etc.) affected performance. There were, however, certain identifiable trends in some data which will be discussed here.

Figure 8-11 shows a relatively precise inverse relationship between volatile solids and iodine number for both spent and regenerated carbon. Only one data point departed significantly from this precise trend; for Run #2 an abnormally high volatile solids was measured and this is not shown on Figure 8-11.

Such a precise relationship between volatile solids and iodine number for spent carbon was not expected. The volatile solids determination used measured both adsorbed organics and any biomass present in the spent carbon. The variable nature of carbon contacting system operation (e.g., number of stages, dosage and SRT) and the length of time of storage of gravity thickened carbon (e.g., 1 to about 20 days) would have been expected to effect the amount of biomass present, and thus prevented determination of such a precise relationship as shown on Figure 8-11.

There is a second definable trend in the data shown on Figure 8-11. It appears that the more exhausted the carbon (i.e., high volatile solids and/or low iodine number), the more difficult high recovery of adsorptive properties was

TABLE 8-7: FBF OPERATING CONDITIONS: VIRGIN CARBON

Run Number	10
Run Date, (1973)	October 20-21
Operation Principal	OGR[a]
Feed Point (from Bed Floor, in.	18
Sand Size, US Mesh	15x30
Gas Flow, SCFM:	
Burner Air	80
Burner Gas	11
Off Gas Recycle	90
Gas Velocity, ft/sec:	
Bottom of Bed	2.6
Freeboard[b]	1.0
Run Length, hr	39.3
Feed Rate, dry lb/hr	17.7
Moisture in Feed, % by wgt	78
Avg Temperature, °F:	
Freeboard	1350
Sand Bed	1450
Firebox	1970
Avg Pressure, in. of H_2O:	
Freeboard	-3
Firebox	63
Avg O_2 Content of Stack Gas, % by volume	0.5

a - off gas recycle
b - includes water vapor

TABLE 8-8: CARBON CHARACTERISTICS: RUN #10 AND VIRGIN CARBON

Run Number	10					
Run Date	October 20-21, 1973		August 2, 1972	September 27, 1972	May 30, 1973	
Fixed Carbon Recovery %	98					
	Feed	Regenerated	Virgin Carbon	Virgin Carbon	Virgin Carbon	Average
Solids, % by wgt						
Volatiles	4.1	1.7	3.0	1.9	3.8	3.2
Ash	3.3	9.6	2.7	3.9	2.8	3.2
Fixed Carbon	92.5	88.7	94.3	94.2	93.4	93.6
Iodine Number	942	1040	1025	980	977	981
Molasses Number	85	83	--	100	--	--
Acid Solubles, % by wgt	1.9	2.2	--	--	--	--
Iron, % by wgt	0.5	1.3	--	--	--	--
Aluminum, % by wgt	0.1	0.1	--	--	--	--

FIGURE 8-10: COMPARATIVE EQUILIBRIUM ADSORPTION ISOTHERM TESTS: REGENERATION RUN NUMBER 10

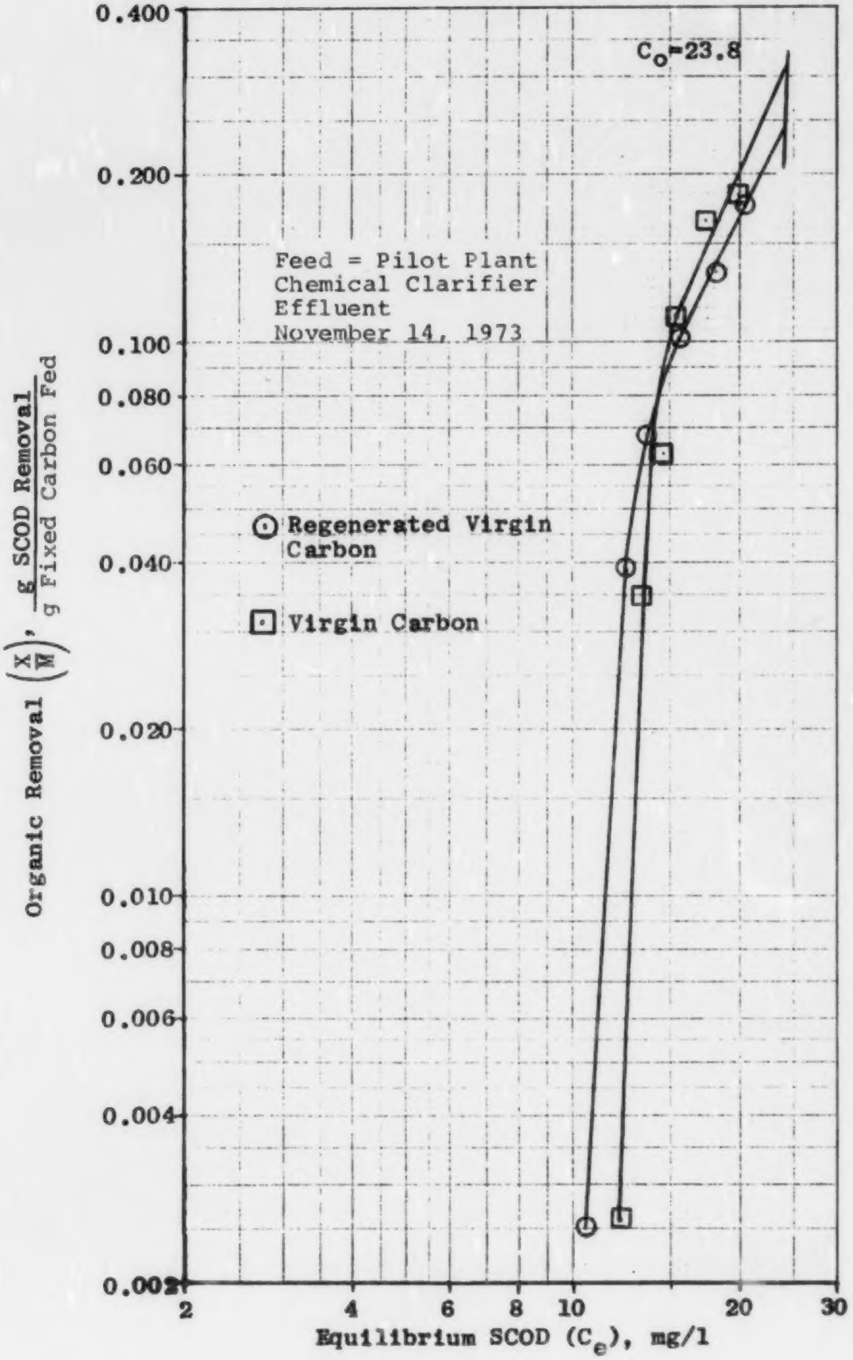
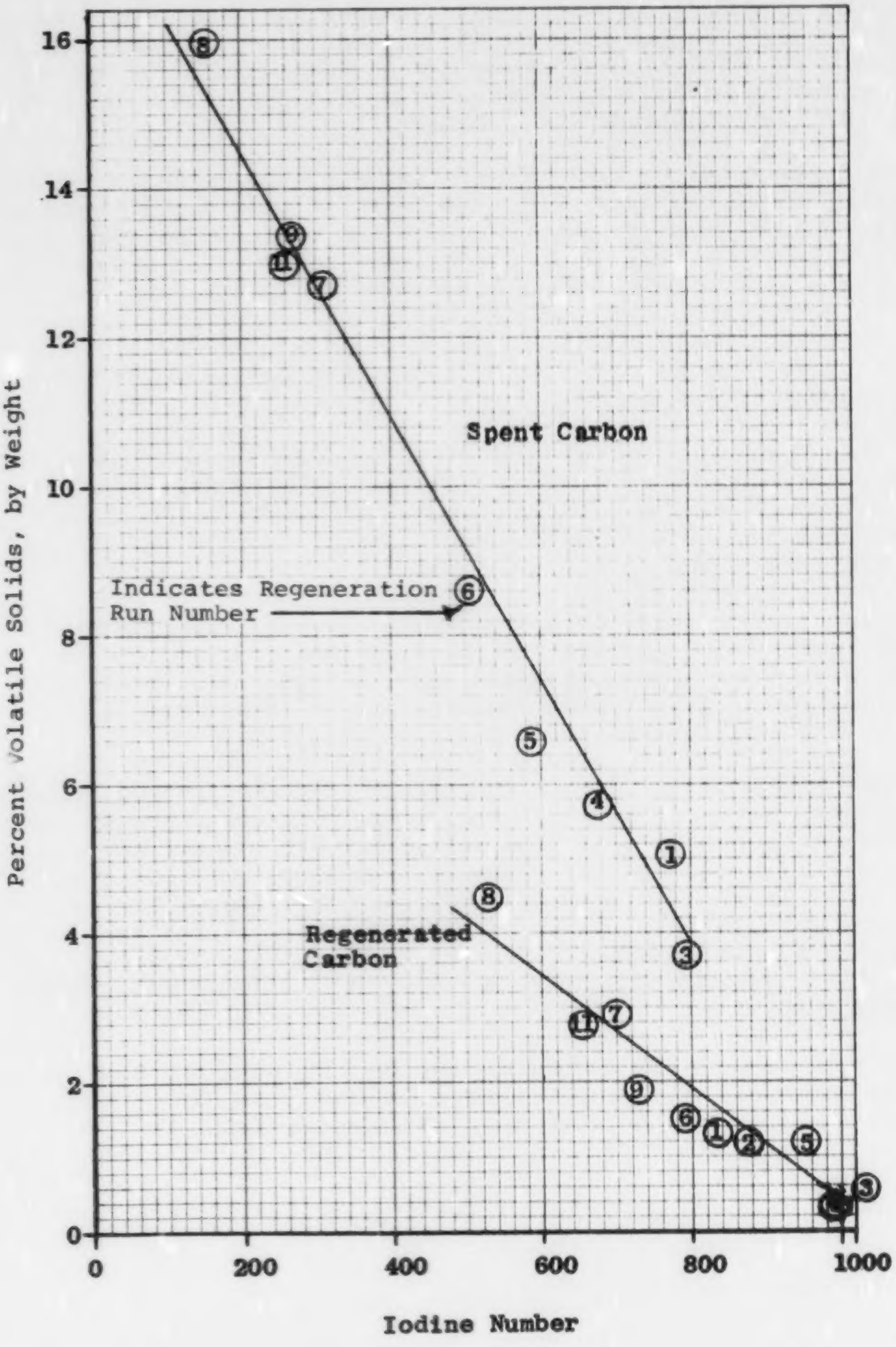


FIGURE 8-11: RELATIONSHIP BETWEEN
IODINE NUMBER AND VOLATILE SOLIDS



to achieve. This observation is seen more clearly in Figures 8-12 and 8-13. Two factors are evident. First, the higher the degree of exhaustion, the higher the incremental recovery of adsorptive property. Secondly, the higher the degree of exhaustion, the further from full recovery of adsorptive properties. The second factor would seem to indicate the existence of a rate limiting phenomenon in the FBF regeneration process.

Any discussion of the significance of the effects of thermal regeneration variables on recovery of adsorptive properties has to be a relative one. There is no precise and accurate universal test for adsorptive properties! In general, the various indirect measurements used in this study indicated that when using the OGR principal of operation, a significant loss of molasses number, iodine number and the ability to adsorb SCOD from chemically treated Salt Lake City wastewater was experienced.

In the final analysis, the effectiveness of regeneration can only be determined by reuse in the PAC-PCT process. As will be shown in Section IX, little difference in SCOD removal could be found when using regenerated carbon in the pilot plant, compared to virgin carbon.

It should be noted that comparative equilibrium adsorption isotherms for regeneration Runs #9 and #11 are not valid comparisons. The comparison should have been made with actual pilot plant feed carbon (i.e., once regenerated carbon) and not virgin carbon.

Using a uniquely different process for carbon regeneration, others have reported that the regeneration product was recovered in a slurry which had several thousand mg/ℓ of SCOD in the bulk solution.[9] During this present study, SCOD's of from 35 to 172 mg/ℓ were found for regenerated carbon slurries. Recycling this SCOD material, which is probably nonadsorbable, would result in adding about one mg/ℓ SCOD to the wastewater. Thus, a process SCOD removal problem would not exist.

The source of SCOD material in the FBF product slurry was undetermined. It could have resulted from poor oxygen and fuel control in the FBF unit causing volatilized, and possibly pyrolized, organics to be scrubbed out of the off-gas.

A major area of concern about the regeneration system performance is the build up of inert, or ash, material upon regeneration and reuse. For the present study, there were three basic sources of inerts:

FIGURE 8-12: RELATIONSHIP BETWEEN
REGENERATED AND SPENT CARBON PERCENT VOLATILE SOLIDS

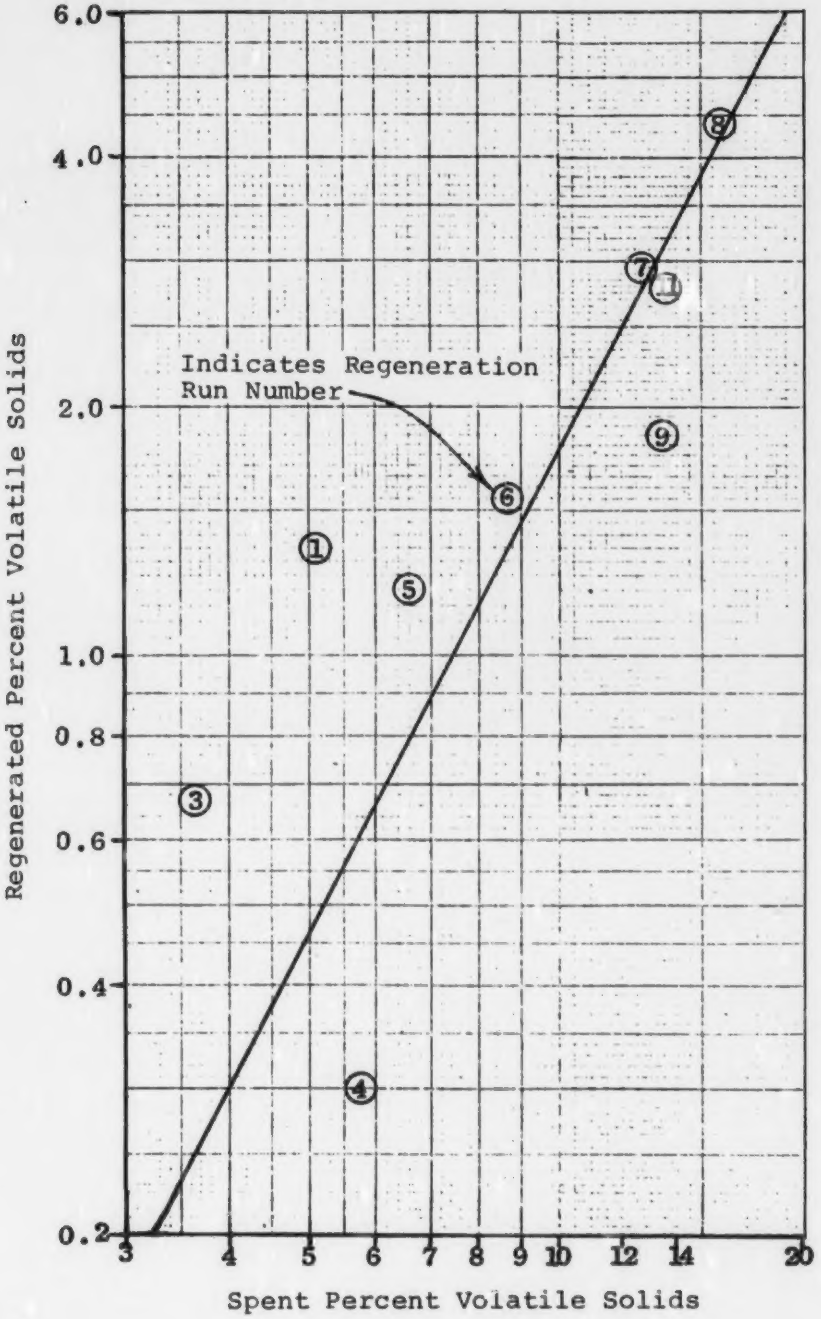
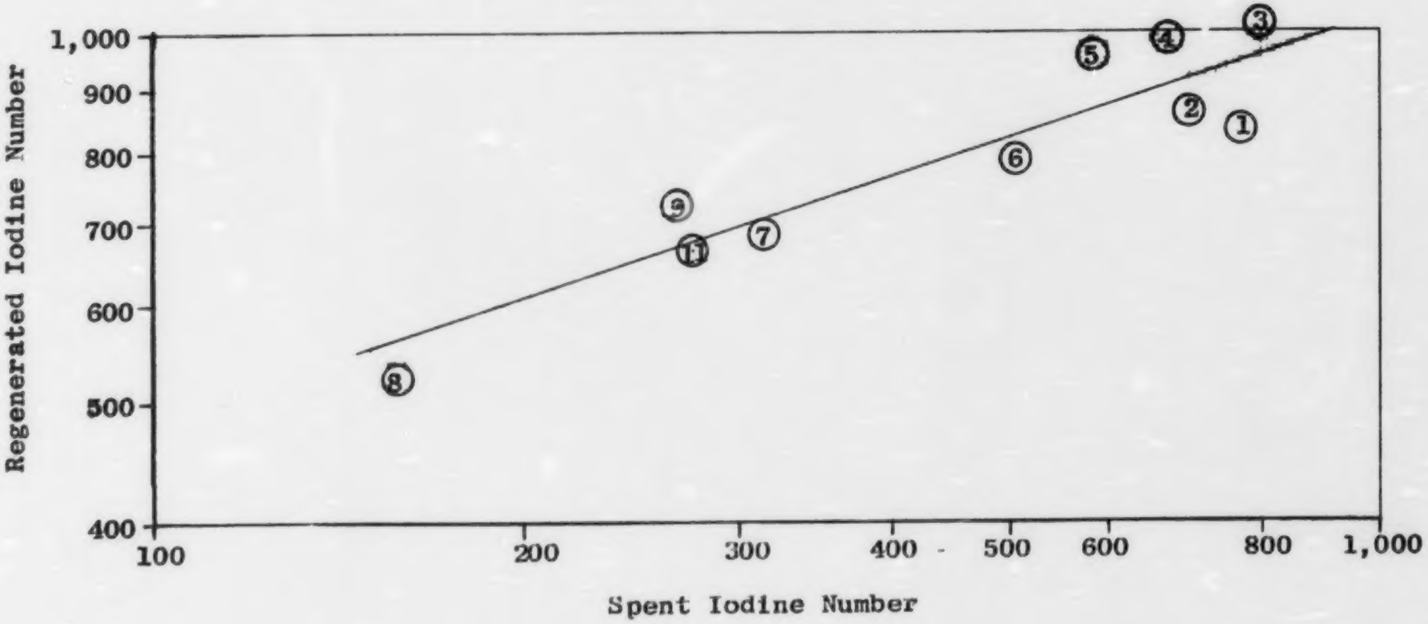


FIGURE 8-13: RELATIONSHIP BETWEEN REGENERATED AND SPENT CARBON IODINE NUMBERS



1. Sand from attrition and/or carry over from the FBF,
2. Chemical precipitates entering the carbon contacting system from the chemical clarifier overflow and
3. Combustion products from the burning of carbon and volatile organics.

Though sand carryover was a problem with the FBF used, it would be hoped that better design of a prototype, including scrubber water screening and/or classification (e.g., hydrocyclone), would minimize this problem.

Essentially all of the chemical compounds, coming from the chemical clarifier overflow floc, present in the FBF product should be acid soluble. In prototype operation, facilities for acid treatment of FBF product should be provided to allow elimination of these chemical inerts.

The amount of inerts generated from combustion of carbon and organics in the FBF should be nominal. For example, if 10 percent fixed carbon losses were experienced, the ash resulting would amount to only about 0.3 lb ash/100 lb of regeneration product. The contribution from combusted organics would be of a similar amount.

In regard to the above discussion which implies little concern over build up of inerts in a carbon regeneration and reuse system, a rather real problem does exist. In Section VI it was indicated that efficient suspended solids removal by the granular media filter resulted in containment and build up of carbon fines in the carbon contacting system. It is probable that a substantial build up of carbon fines will result from attrition of carbon in the carbon contacting and regeneration operations. In addition, normal attrition of sand and refractory in the FBF will probably add to the fines to be handled in the carbon contacting and clarification system. The best approach to removal of fines from the carbon system would be to, when necssary for control, recycle granular media filter backwash to the chemical clarifier. The exact amount of blowdown of fines required is not precisely definable, but it should be no more than a few percent of the carbon used.

Evaluation of FBF capacity, lb dry feed solids/hr, results for this study indicates widely varying data. For example, when employing the BIG operating principal (Runs #1 thru #5), average feed rates ranged from 35 to 71 lb/hr. When employing the OGR operating principal (Runs #6 thru #11), average feed rates ranged from 13 to 37 lb/hr. The FBF feed rate capacity is related to heat input (combustion gas), heat losses (in off-gas and radiation), feed moisture content, thermal efficiencies and other factors. The near random variation of these parameters during this study precluded definition of cause-effect relationships with FBF feed rate.

An FBF operating factor was determined which was found to be precisely related to FBF feed rate. The following rationale was used to define this "FBF Operating Factor". Feed rate should be directly proportional to heat input which can be represented by combustion gas flow rate. Part of the thermal efficiency of the unit is related to the amount of gas which must be heated to operating temperature. Thus, feed rate should be inversely proportional to the sum of combustion air and off-gas recycle flow rates. The feed rate should be directly related to heat consumed. It can be shown that the heat consumed is directly related to the difference in firebox and sand bed temperatures, and other factors.

Based on the above rationale the following "FBF Operating Factor" was computed for each regeneration run:

$$FBF_{of} = \frac{(T_{FB} - T_B)\ Q_g}{Q_a + Q_{ogr}}$$

Where:

Q_g = combustion gas flow, SCFM

Q_a = combustion air flow, SCFM

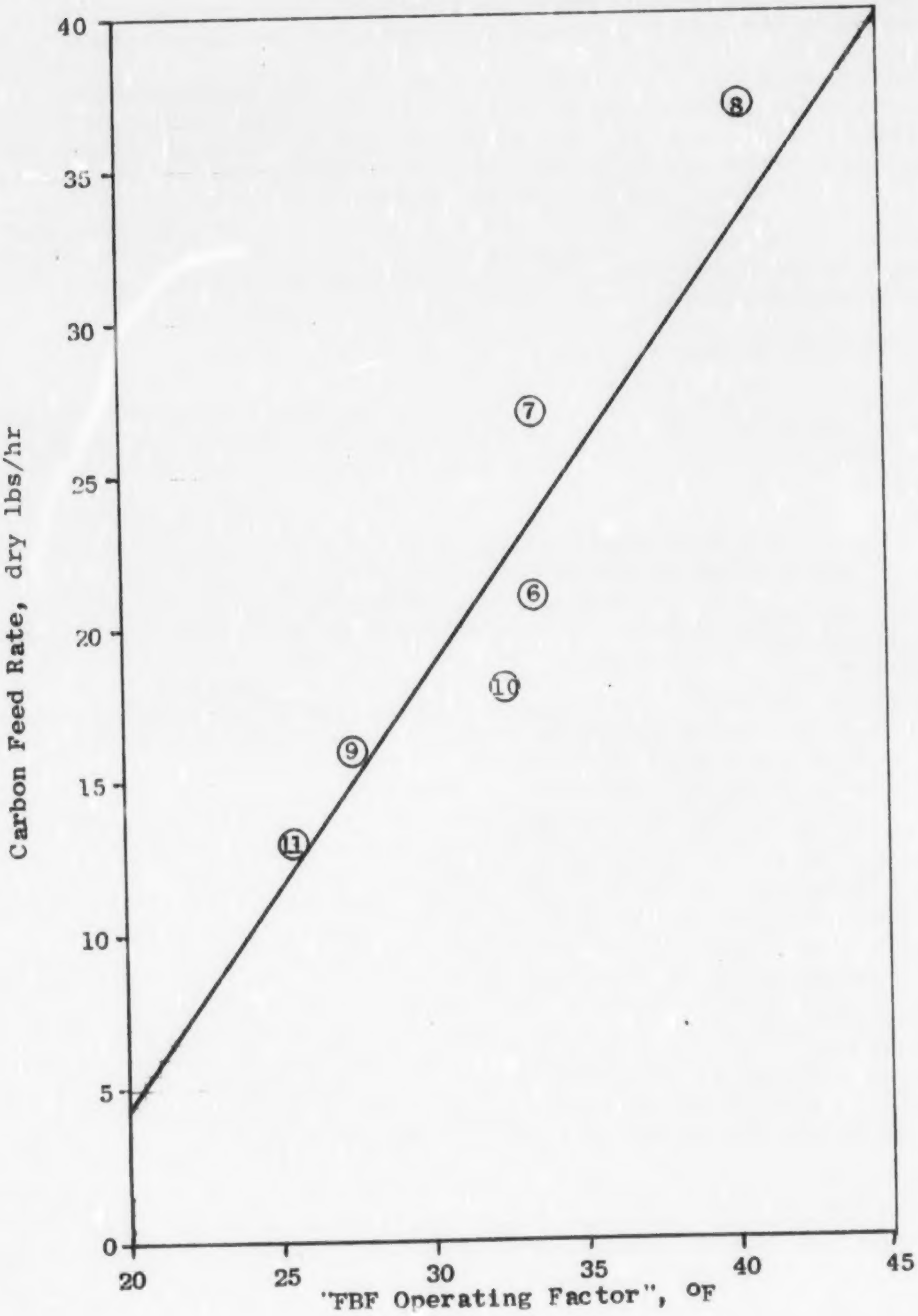
Q_{ogr} = off-gas recycle gas flow, SCFM

T_{FB} = firebox temperature, °F

T_B = sand bed temperature, °F

A plot of FBF feed rate versus this "Operating Factor" is shown on Figure 8-14. A linear regression correlation coefficient of 0.88 was computed, indicating a fairly precise fit of the data. A plot including the BIG results would show approximately the same relationship and precision. It is not the intent of the authors to imply a general linear relationship, but rather only for the range of data shown. It should also be noted that the parameters used to compute the "Operating Factor" are not mutually exclusive variables. The data in Figure 8-14 implies that to maximize FBF capacity,

FIGURE 8-14: FURNACE FEED RATE
VERSUS FBF OPERATING FACTOR



maximizing heat input (Q_g), minimizing off-gas recycle (i.e., operate at highest practical firebox temperature) and the lowest practical sand bed temperature is necessary. The choice of sand bed temperature would have to take into account the effectiveness of regeneration obtainable.

In summary, it is concluded that the FBF regeneration system using the OGR principal of operation efficiently regenerated carbon. Fixed carbon recoveries of 76 to 100 percent were experienced with an overall average in excess of 90 percent. Some loss of carbon adsorptive properties was experienced. The limited time during which efficient recoveries were experienced precluded being able to identify operating conditions which would maximize recovery of adsorptive properties for the system studied.

OPERATING PROBLEMS

During the course of this study numerous operating problems were experienced. Not all problems affected results presented and discussed, but they are considered worthy of discussion.

Perhaps the most frustrating problem, because of its undefinable nature, was short-circuiting of firebox gases around the tuyere arch (a castable refractory member). Short-circuiting was indicated by the existence of hot spots running from the firebox to the sand bed regions of the steel shell. Several probable causes of short-circuiting were considered. Improper installation of refractory materials was a prime suspect. Probably just as significant was numerous heat-up and cool-down cycles and various modifications made to the FBF unit. Several new carbon feed points were jack hammered through the sand bed zone refractory wall. Six holes were also jack hammered through the tuyere arch. The existence of firebox gas short-circuiting did not seriously effect performance when using the OGR operation principal. Under this condition, firebox gas presumably contained only nominal amounts of oxygen (e.g., less than 0.5 percent by volume). However, when using the BIG operating principal, firebox gas contained as much as 12 to 17 percent oxygen by volume. Short-circuiting of gas could have resulted in carbon being exposed to oxygen at ignition temperatures at which it would have been burned. It is unlikely that short-circuiting was so extensive as to have caused the 40 to 60 percent losses of carbon experienced.

The second major operational problem related to the measurement of oxygen. A hand Orsat unit, with a 0 to 7 percent oxygen by volume scale, was routinely used. The device was considered reasonably accurate but somewhat imprecise in the FBF operating range of less than one percent oxygen by volume. Attempts were made to use a continuously sensing and indicating device. The need for near continuous calibrations and erratic performance resulted in the device being abandoned.

A more serious problem existed due to the location of sampling for oxygen measurements. The sample point used was at the off-gas stack, after the venturi scrubbers. This location was necessary when using the BIG operating principal. Adjustment of air and gas flow rates prior to carbon feeding allowed achievement of a desirable net exit oxygen level. Whether or not this desired oxygen level existed in the FBF freeboard or upper reaches of the sand bed as well as in the off-gas stack was not practically determinable.

For OGR operation, the off-gas stack sample point was, after the fact, determined to not be the most desirable. Prior to feeding carbon, measurement of oxygen at this location was appropriate. However, if adjustments in combustion gas or air flow rates were required during carbon feeding, the net effect of these adjustments on firebox oxygen content was not determined. It would have been more appropriate to have monitored the oxygen present in the firebox gas.

Another annoying problem experienced was plugging of the FBF carbon feed line. This line was originally installed as a 2-inch diameter pipe. At 60 lb/hr dry carbon feed rate, the velocity was about 2.9 fpm. During some runs, this line plugged two or three times. Plugging necessitated reduction of fluidizing gas flow to below minimum fluidization flow, dismantling of the feed line plumbing and rodding of the carbon into the hot sand bed.

After the seventh regeneration run, the two inch diameter feed line was replaced by a 1-inch diameter line with long radius elbows. The four fold increase in velocity reduced feed line plugging problems substantially.

Another carbon feed related problem existed in the vacuum filter cake hopper. A ribbon screw mixed the cake, forming a paste, which was readily pumped by a Moyno pump. Filter cake frequently bridged the six inch throat, just above the ribbon screw, resulting in no carbon reaching the pump. Interruption of FBF feed due to bridging resulted in rapid increase in sand bed temperature. The bridging tendency in the filter cake hopper necessitated frequent operator attention.

Another problem frequently encountered was back flow of sand into the 6-1/4-inch diameter BIG nozzles. After Run #3 individual gas flow meters were installed on each nozzle. This allowed routine monitoring of gas flow and thus minimized poor distribution of BIG.

On occasion, small quantities of sand were found on the firebox floor. This resulted from back flow of sand through the tuyere orifices. The extent of back flow was considered nominal and not a significant problem.

A previous study using a similar FBF unit with OGR operation indicated extensive corrosion problems in the OGR system.[2] Based on their experience, OGR plumbing used in this study was reinforced fiberglass piping. No corrosion problems were experienced. The type of OGR blower used in the current study had mild steel rotors. These rotors were severely corroded after about two regeneration runs and necessitated replacement of the blower.

COMMENTS ON "BIG" OPERATING PRINCIPAL

As previously noted, the authors concluded that the BIG operating principal with the FBF unit used in this study did not result in acceptable recovery of fixed carbon. The fundamental problem was defined as being non-preferential combustion of natural gas in the lower portion of the sand bed, prior to the carbon feed point. There are at least two factors which might explain why this problem was insurmountable in this study.

The first factor relates to the comparative flame propagation rates of carbon and the natural gas fuel used. Flame propagation rate is the velocity at which a flame will travel in a fuel air mixture. Data on the flame propagation rate for the carbon used in this study could not be found. It was determined that the flame propagation rate for pulverized soft coal was considerably in excess of that for natural gas.[10] It is thus inferred that, were carbon and natural gas (BIG) to have been simultaneously exposed to localized excess oxygen, that the carbon would have been preferentially burned to a higher degree than would the natural gas. That localized pockets of gas with excess oxygen might have existed in the sand bed is entirely conceivable when considering the second factor to be discussed; namely, the quality of sand bed fluidization. A large body of published knowledge exists concerning fluid-bed technology. Several observations were made during operation of the pilot plant FBF unit with the BIG operating principal which might allow bringing to bear some of this technology in an attempt to understand why unsuccessful carbon recovery results were experienced.

Almost without exception the pressure drop across the bottom 18-inches of the fluidized sand bed fluctuated widely during operation. The absolute value of pressure drop varied with fluidization velocity, bed temperature and size of sand, but the fluctuations were consistently about ±6 inches of water. When visually observing the fluidized bed, through a sight glass mounted on top of the unit, it was also observed that, almost without exception, the top of the bed would vacilate over about a ft depth. Typically the top of the bed would stay at about the 4-ft marker for several seconds and then suddenly, and somewhat violently, surge to the 5-ft marker and stay there for about a second. The entire FBF unit could be felt to pulse or shudder when this occurred. Some sand grains were thrust upward, well into the freeboard zone during these bed surges. These observations are indicative of a "slugging" fluidized bed.[11] Slugging occurs due to the growth of rising gas bubbles when experiencing a uniformly reduced pressure region and also partly due to coalescence of bubbles. Slugging typically occurs when fluidization velocity is considerably greater than the minimum particle fluidization velocity.

The type of tuyere system used could also have had an effect on slugging. Ideally, the greater the number of uniformly spaced orifices and higher the pressure drop, the more uniform the distribution of gas flow and the smaller the gas bubbles formed at the orifice outlet. Presumably it was this factor that prompted the recommendation of installation of additional tuyeres to improve the quality of bed fluidization and mixing.

A second set of observations related to the presence of fuel (natural gas) in the FBF freeboard zone. With the FBF at operating temperatures and levels of fluidization, bursting of bubbles was observed at the surface of the sand bed. Obviously, the bubbles were not seen, but rather the motion and color of sand at the bed surface. When a bubble would burst at the bed surface, the color of sand would glow more intensely in that region. Bubbles would break the bed surface forming 6 to 14-inch diameter circles of sand discoloration. As a bubble would break at the bed surface, a flash of light usually occurred at that spot. This obviously indicated ignition of natural gas. Whether the bubbles or the freeboard under certain operating conditions was documented by temperature profile print out.

The observation of "bursting bubbles" and burning of natural gas in the freeboard zone substantiated that slugging was occurring and that the quality of fluidization was typical of gas fluidized sand beds.[11]

The reasons for slugging being generally observed are not clearly understood by the authors. As previously noted, normal fluidization velocities especially at the bottom of the variable area bed container, were considerably in excess of minimum fluidization velocities. This condition typically results in slugging.[11] In addition, the relatively large bed depth to diameter ratio would have exaggerated slugging. Another factor which might have promoted slugging was the relatively low designed tuyere headloss of 10-inches of water. Others suggest that a minimum of 14-inches of water be used for design.[11]

During carbon feeding, a red haze of carbon existed in the freeboard zone making it impossible to view the surface of the fluidized bed. Considering that carbon, with it's associated water, was fed at a single point and that the vaporized water volume was as high as 50 percent of firebox gas flow, it is quite probable that gross slugging occurred just above the carbon feed point. This factor plus the typical non-random nature of gas fluidized sand beds might have resulted in gross short-circuiting of carbon to the surface of the sand bed. It is probable that carbon short-circuiting resulted in residence times much shorter than indicated by average gas velocities and depth of travel.

Another factor which might have caused short-circuiting in the freeboard zone and thus promoted slugging was the relatively shallow freeboard depth. As can be seen in Figure 8-2 when the sand bed was fluidized to the 5-ft level, there was only about a 4-ft freeboard.

Based on the authors' experience presented in this report and their interpretation of a limited amount of published literature on the subject of fluidized bed processes, they seriously doubt that the BIG operating principal could be made to work in a prototype carbon FBF unit. There are, however, several suggestions to be made which substantially increase the probability of success.

The bed injection fuel used should have a flame propagation rate well in excess of carbon if such a fuel is economically available. Perhaps the operating fluidization velocity should be limited to minimize slugging. Much reduced bed depth to diameter ratios for prototype units should also reduce the tendency for slugging. Deeper freeboard would also be suggested to reduce any tendency for slugging and also to minimize sand carryover. Perhaps increased numbers of tuyeres with smaller orifices and higher pressure drops would also reduce the slugging tendency. It is possible that a special tuyere could be designed which could accomplish more efficient mixing of firebox gas (with its excess oxygen) and bed injection fuel.

Considering the substantially higher capacity per unit area when using the BIG operating principal it would be hoped that those truly skilled in the art of fluidized bed processes will actively pursue further development in this area.

SECTION IX

CARBON SYSTEM RESPONSE

The singular purpose of carbon treatment was removal of soluble organics from the wastewater. To describe the carbon system and to form a basis for comparing treatment effects, a model of the carbon system response is desirable. This response model should interrelate, at least, carbon system feed and effluent soluble organics and carbon dosage. The identification of a carbon system response model was not an original objective of this study.

Periods of reasonably stable carbon system operation and performance were identified before pertinent system variables could be quantified. The first step was to plot effluent quality, carbon dosage and carbon system solids retention time (SRT) for each day of pilot plant operation. Routinely collected solids-contact unit solids profile data indicated uniform solids concentration in the carbon contactor units. Therefore an approximate carbon SRT was calculated as follows.[1]

$$\text{SRT} = \frac{\text{(Reaction Zone Solids) (Volume of Carbon Slurry)}}{\text{(Flow Rate) (Carbon Dosage)}}$$

The numerator is an approximation of the inventory of carbon (pounds) within the solids-contact units and the denominator is the carbon fed (lb/day).

Various treatment campaigns were determined on the basis of staging, chemical pretreatment, carbon dosage, effluent quality, SRT and use of regenerated or virgin carbon. SCOD was used as the measure of organic material.

Tables 9-1 and 9-2 present average SCOD, SRT and carbon dosage data. Reduced data shown indicate incremental organic reductions and organic removals for the various campaigns. During pilot plant operation, granular media filtration of all carbon contactor effluent was impossible (e.g., when carbon contactors were run in parallel). Thus, the effluent

TABLE 9-1: CARBON TREATMENT SYSTEM: SUMMARY OF SINGLE-STAGE DATA

Campaign #	Observed Data				Reduced Data			Run Length days
	Fixed Carbon Dosage (M), mg/ℓ	SCOD, mg/ℓ			SCOD Removed, mg/ℓ	Organic Removal, g/g	Carbon SRT, days	
		Raw Wastewater (C_r)	Carbon System Feed (C_o)	Effluent (C_e)	All Stages (X)	All Stages (X/M)		
VL1-1[a]	319	28.8	36.9	8.0	28.9	0.091	2.8	53
VL1-2	298	26.8	35.4	7.6	27.8	0.093	2.8	45
VL1-3	0	42.3	30.9	15.0	15.9	---	∞	7
VL1-4	98	42.3	30.9	7.0	23.9	0.244	6.5	7
VL1-5	0	46.0	45.7	34.0	11.7	---	∞	23
VL1-6	31	46.0	45.7	31.5	14.2	0.458	14	23
VL1-7	198	40.3	38.6	10.2	28.4	0.143	5.9	34
VL1-7'	141	35.8	36.3	8.8	27.5	0.195	7.4	16
VL1-8	141	35.8	36.3	8.7	27.6	0.196	6.8	16

[a] VL1-1 = Virgin Carbon; Lime Pretreatment; 1 Single-Stage Carbon Treatment; 1-Run #1

TABLE 9-2: CARBON TREATMENT SYSTEM: SUMMARY OF TWO-STAGE COUNTER-CURRENT DATA

Campaign #	Observed Data					Reduced Data					Carbon SRT, days	Run Length days
	Fixed Carbon Dosage (M), mg/ℓ	SCOD, mg/ℓ				SCOD Removed, mg/ℓ			Organic Removal g/g			
		Raw Wastewater (C_r)	Carbon System Feed (C_o)	Intermediate Feed (C_i)	Effluent (C_e)	1st Stage (X")	2nd Stage (X')	All Stages (X)	All Stages (X/M)	2nd Stage (X'/M)		
VA2-9[a]	106	39.0	23.8	10.0	4.2	13.8	5.8	19.6	0.185	0.055	14	13
VA2-10	294	40.1	23.0	7.0	2.0	16.0	5.0	21.0	0.071	0.017	6.1	19
VA2-11	103	40.7	22.7	13.1	5.0	9.6	8.1	17.7	0.172	0.079	15	32
VF2-12	108	34.3	19.6	8.7	2.7	10.9	6.0	16.9	0.156	0.056	9.8	49
VF2-13	75	30.3	17.3	6.6	3.0	10.7	3.6	14.3	0.191	0.048	20	21
RF2-14	101	32.5	17.9	7.1	3.8	10.8	3.3	14.1	0.140	0.033	13	23
RA2-15	72	42.8	24.0	17.4	7.0	6.6	10.4	17.0	0.236	0.144	18	34
RA2-16	28	47.6	24.3	15.9	10.6	8.4	5.3	13.7	0.489	0.189	30	7

[a] VA2-9 = Virgin Carbon; Alum Pretreatment; 2 Two-Stage Counter-Current Operation; 9-Run #9

SCOD concentrations shown in Table 9-1 are from the final carbon contactor and not the granular media filter. This approach is consistent with previously reported results.[1] With few exceptions, the effluent SCOD's of the granular media filter and carbon contactor effluents were equal.

The campaign numbers in Table 9-1 and 9-2 were coded to indicate certain operational variables. The first letter indicates the use of virgin (V) or regenerated (R) carbon. The second letter indicates chemical pretreatment, lime (L), ferric chloride (F) or alum (A). The first number indicates single-stage (1) or two-stage counter-current (2) operation. The number after the hyphen indicates the chronological campaign number. For example, Campaign VA2-10 is virgin carbon, alum pretreatment, two-stage counter-current contacting and campaign number 10.

CARBON SYSTEM RESPONSE MODEL

Several types of models have been used to describe real time powdered carbon-organic removal system responses.[1,12,13] In a previous report the authors used both the classical Freundlich adsorption model and an empirical model as follows:[1]

$$\frac{X}{M} = a \times b \frac{C_o}{M}$$

Where:

X = organics removed (C_o-C_e), mg/ℓ

M = fixed powdered carbon dosage, mg/ℓ

C_o = system feed organics, mg/ℓ

C_e = system effluent organics, mg/ℓ

a,b = emperical constants

This emperical model, X/M vs C_o/M, was first used by others.[12]

Correlation of system feed and effluent SCOD's and carbon dosages was found to be very imprecise when using the Freundlich model; whereas, use of the X/M vs C_o/M model resulted in very precise data correlations. Because of the variability of the powdered carbon system performance results, the use of neither model allowed determining statistically significant treatment effects, even though the actual data suggested substantial effects due to, for example, the number of counter-current contacting stages.[1]

Both of the above models have unreal extrapolated boundry conditions. The X/M vs C_o/M model predicts the unreal condition of being able to achieve zero effluent SCOD at a large, yet finite, carbon dosage. Extrapolation of the Freundlich model indicates an asymptotic approach to zero effluent SCOD at infinitely large carbon dosages.

Extrapolation of the Freundlich model to very low carbon dosage, approaching zero, predicts that the carbon system effluent SCOD will approach the influent SCOD. On the other hand, extrapolation of the X/M vs C_o/M model to zero carbon dosage predicts that the system effluent SCOD will be significantly less than the system feed SCOD. As will be shown later in this Section this latter boundry condition was observed for the powdered carbon system used.

Because of the factors discussed above, there was no clear basis for choosing one or the other model to evaluate the results obtained during this study. Due, in part, to the following reasons, the decision was made to use the Freundlich model.

1. Most published articles dealing with the application of powdered carbon treatment of municipal wastewaters have used the Freundlich model.

2. Laboratory equilibrium adsorption responses are usually precisely correlated by the Freundlich model.

3. Use of the X/M vs C_o/M model provides no new insight into the evaluation of treatment effects on the powdered carbon system response.

Figure 9-1 is a Freundlich plot of organic removal data for single and two-stage counter-current operation using virgin carbon. The lines shown are linear regressions of log X/M on log C_e. Table 9-3 presents several such regression equations for organic removal data in Table 9-1 and 9-2.

It is apparent from the poor correlation coefficient in Table 9-3 and the relative slope in Figure 9-1 that the single-stage curve does not realistically describe the system response. The slope is much too flat relative to the laboratory equilibrium adsorption isotherms and to the two-stage results. In general, it appears that organic removals, X/M, for two-stage counter-current treatment are generally higher than for single-stage treatment.

FIGURE 9-1: SCOD REMOVAL-ADSORPTION RESPONSE: PILOT PLANT RESULTS USING VIRGIN CARBON

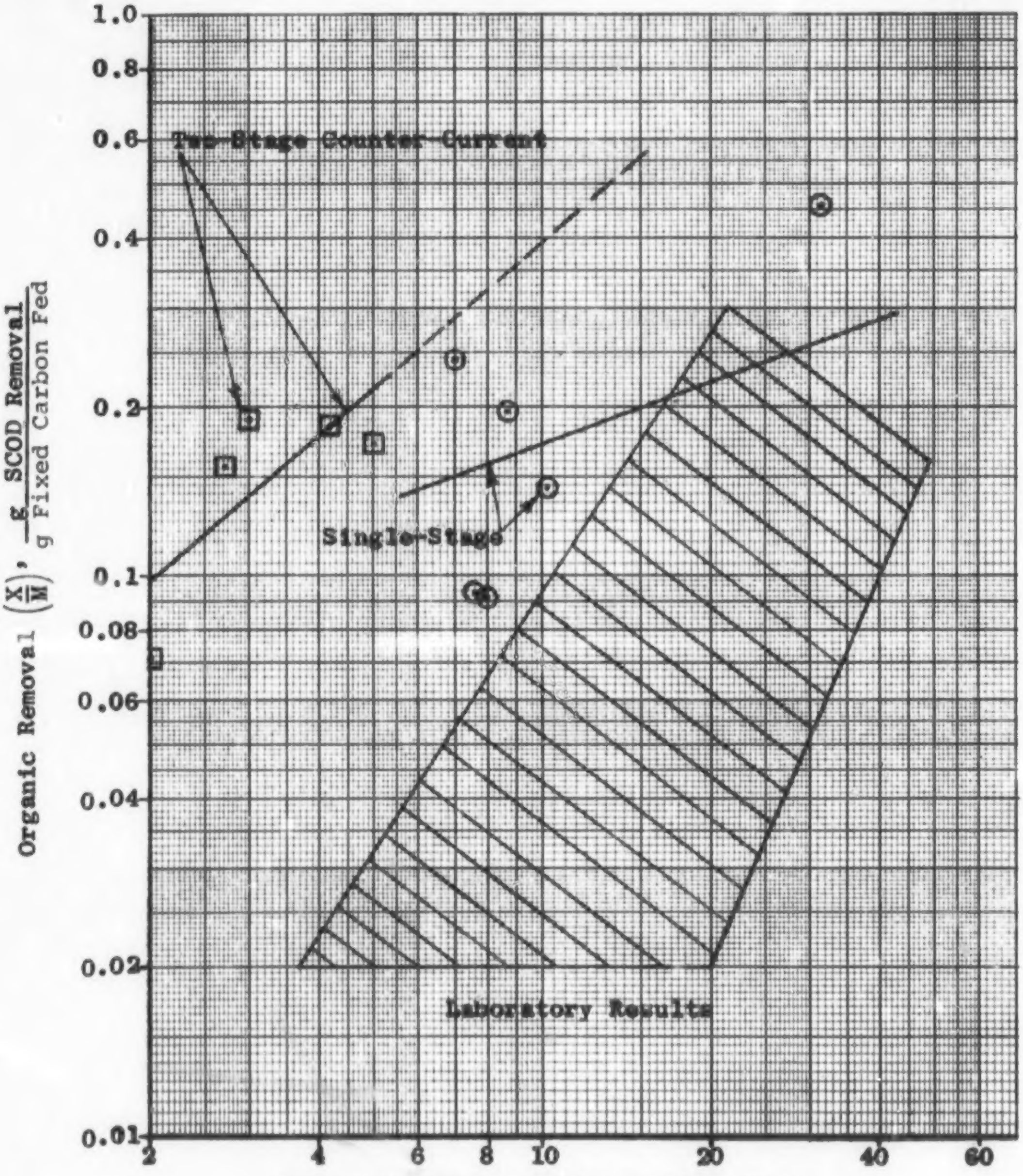


TABLE 9-3

REGRESSION ANALYSIS LOG X/M ON LOG C_e

Contact Stage	Regression Equation	Correlation Coefficient
Single Stage (Virgin Carbon)	$\frac{X}{M} = 0.073\ C_e^{0.37}$	0.50
Two-Stage Counter-Current (Virgin Carbon)	$\frac{X}{M} = 0.053\ C_e^{0.87}$	0.76
Two-Stage Counter-Current (Virgin & Regenerated Carbon)	$\frac{X}{M} = 0.049\ C_e^{0.90}$	0.90
First Stage of 2SCC[a] (Virgin & Regenerated Carbon)	$\frac{X}{M} = 0.018\ C_e^{0.99}$	0.71
Second Stage of 2SCC (Virgin & Regenerated Carbon)	$\frac{X}{M} = 0.009\ C_e^{1.31}$	0.91
First Stage & Second Stage (Virgin & Regenerated Carbon)	$\frac{X}{M} = 0.010\ C_e^{1.23}$	0.92

[a] 2SCC = Two-Stage Counter-Current

The cross-hatched area in Figure 9-1 represents the range of laboratory equilibrium adsorption isotherm test results obtained during this study. Organic removal during these tests was presumably by an adsorption mechanism only. Organic removal in the pilot plant carbon system is seen to be considerably higher than predicted for adsorption by the laboratory adsorption test results. This fact reinforces the thesis that a removal mechanism (biological organic removal) in addition to adsorption existed in the carbon contactor.

Figure 9-2 is a plot of all single-stage and two-stage counter-current data from a previous study and the present study with regression curves.[1] The purpose for presenting all available data was to increase the range and hopefully the precision of the data correlation. In a general sense the response shown is realistic because it does show the expected difference between single and two-stage systems. At low C_e's, the difference in X/M between single and two-stage systems is large. As C_e becomes large, the differences in X/M becomes increasingly smaller.

It is unrealistic to believe that only an adsorption mechanism is operative in the system studied. The differences in pilot plant and laboratory organic removal can best be explained as due to the presence of anaerobic biological activity. The impact of any other treatment effects on the carbon system response is not evident in Figures 9-1 and 9-2. The following subsection is an attempt to elucidate various treatment effects.

ANALYSIS OF TREATMENT EFFECTS

Number of Stages

Figure 9-3 is a plot of fixed carbon dosage (M) versus effluent SCOD (C_e) for all pilot plant data. The regression curves shown (see Table 9-3) were calculated using an average feed of 37.3 mg/ℓ SCOD for the single-stage results and 21.6 mg/ℓ SCOD for the two-stage counter-current contacting results.

At face value the results shown in Figure 9-3 suggest that the number of stages used (two versus one) resulted in a different system response. At virgin carbon dosages of 75 mg/ℓ or more, the average system effluent contained 3.4 mg/ℓ SCOD (standard deviation of 1.2) for two-stage counter-current contacting. For single-stage contacting, at carbon dosages of 98 mg/ℓ or more, the system effluent

FIGURE 9-2: SCOD REMOVAL-ADSORPTION RESPONSE: COMBINED RESULTS OF CURRENT AND PREVIOUS STUDIES

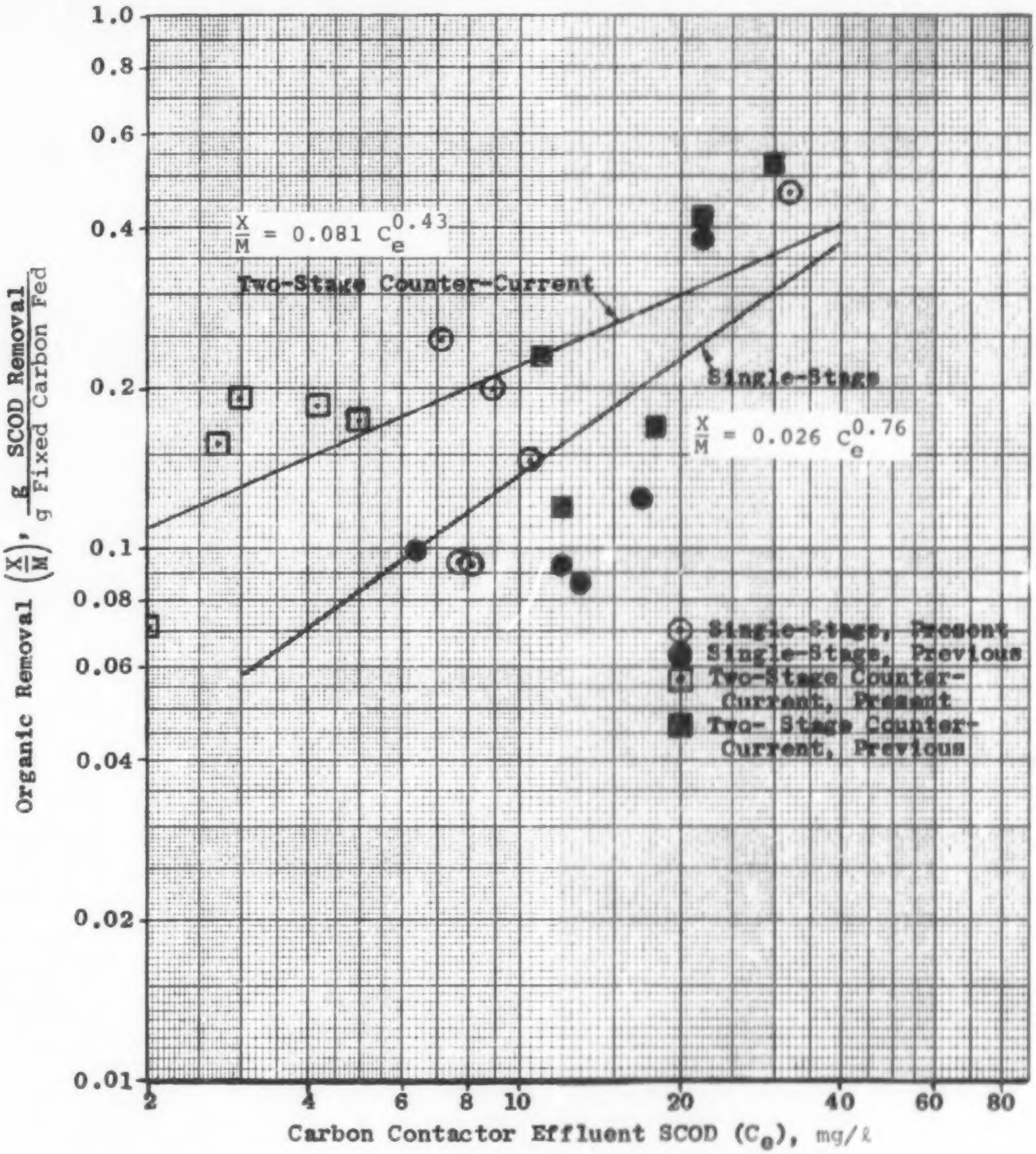
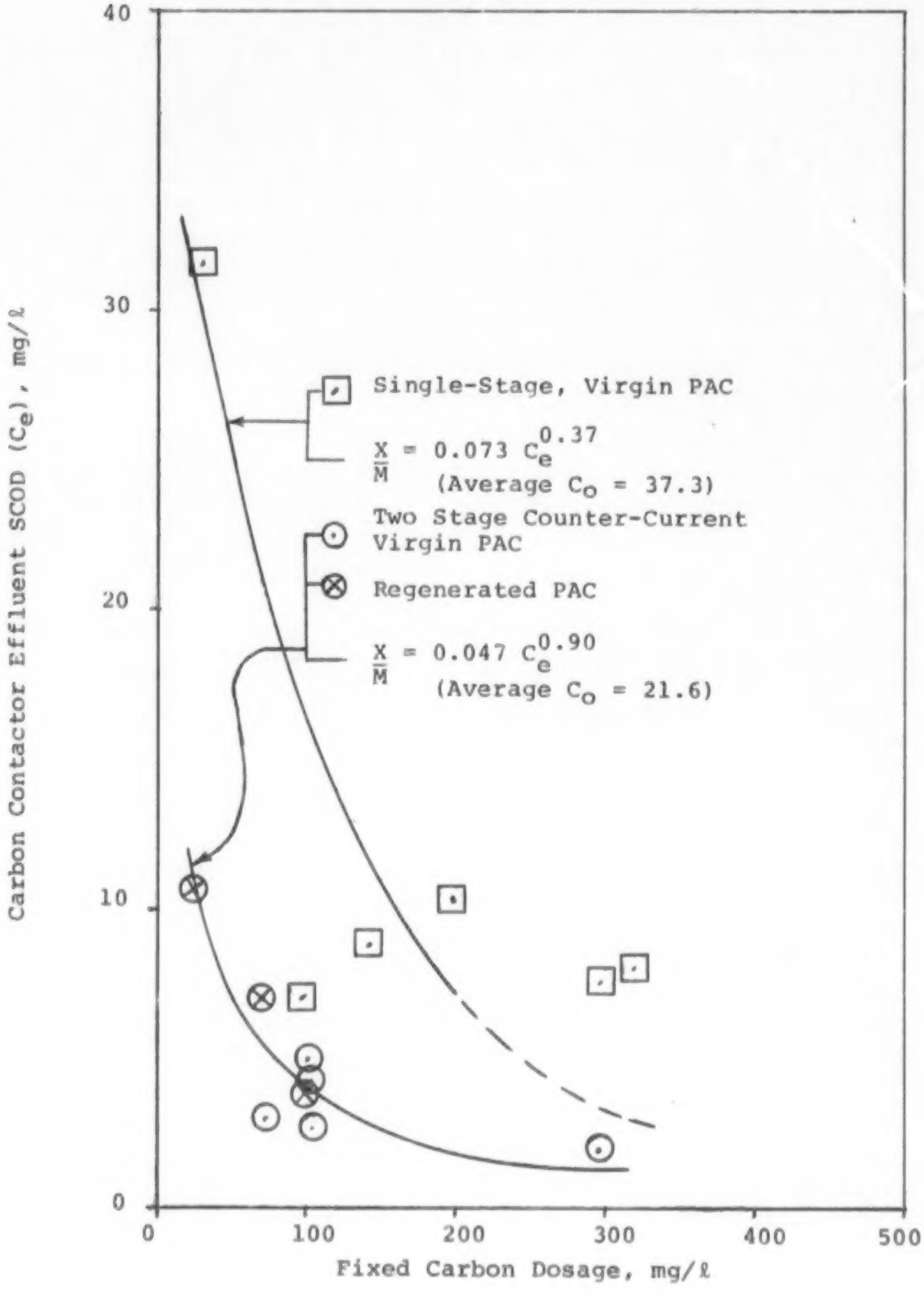


FIGURE 9-3:
SCOD REMOVAL - ADSORPTION RESPONSE:
PILOT PLANT EFFLUENT SCOD RESPONSE TO CARBON DOSAGE



contained an average of 8.4 mg/ℓ SCOD (standard deviation of 1.1). These average effluent SCOD concentrations are statistically significantly different at a probability level of 0.99.

The fact that average carbon system feed SCOD differed substantially for the two contacting modes (37.3 vs 21.6 mg/ℓ) would tend to negate the validity of any inference that the system responses shown in Figure 9-3 were effected by the number of contacting stages.

Evaluation of SCOD removal data in Tables 9-1 and 9-2 indicates that for virgin carbon dosages greater than 75 mg/ℓ the single-stage contacting mode removed 77 percent of the SCOD (standard deviation of 1.8). The two-stage countercurrent system removed from 78 to 91 percent SCOD with an average of 84 percent removal (standard deviation of 5.0). The use of two-stages of contacting resulted in only slightly improved percent removal of SCOD.

It must be concluded that the results of this study do not allow quantification of the effect of number of contacting stages on the powdered carbon-organic removal system response.

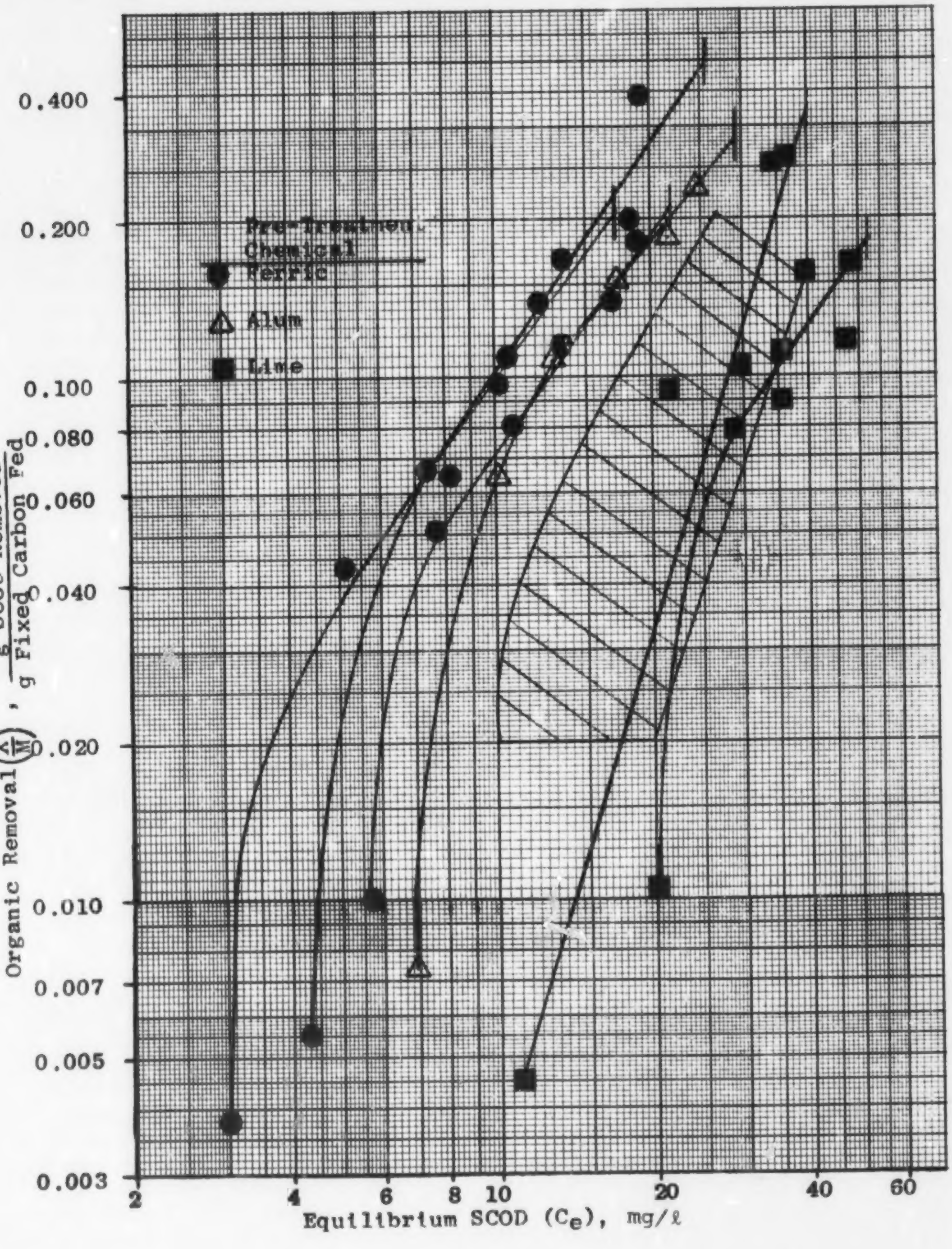
Chemical Pretreatment Effects

On Figure 9-3 there is an obvious and statistically significant difference in system feed SCOD concentration between single and two-stage results. This difference can be related to chemical pretreatment effects. All single-stage operation was done with hydrated lime pretreatment and all two-stage operation was done with inorganic coagulants.

In a previous report (reference 1) it was concluded that there was no significant effect on organic removal by different chemical pretreatments. This conclusion was made on the basis of laboratory work only. When actual pilot plant results from the current study are evaluated, the effect of chemical pretreatment becomes apparent.

Figure 9-4 is a plot of laboratory equilibrium adsorption isotherm tests run on the effluent from the chemical treatment unit. The cross-hatched area represents the range of laboratory adsorption equilibrium isotherm tests run on untreated raw wastewater during previous studies.[1] The data show that an apparent unadsorbable fraction was reduced from 10 to 20 mg/ℓ SCOD for lime or no treatment to 3 to 7 mg/ℓ SCOD with alum or $FeCl_3$ pretreatment. Figures 9-3 and 9-4 show that the feed SCOD concentrations produced

FIGURE 9-4: SCOD REMOVAL—ADSORPTION RESPONSE: LABORATORY EQUILIBRIUM ADSORPTION ISOTHERMS



by alum or $FeCl_3$ pretreatment are considerably lower than SCOD feed concentrations produced by lime pretreatment. It can be inferred from data on Figure 9-3 that, because of the removal of apparently unadsorbable material, lower effluent SCOD (C_e) was achieved with the inorganic coagulants.

Previously reported laboratory data indicated that both high pH lime and inorganic coagulant pretreatment significantly reduced SCOD; however, equilibrium adsorption isotherm responses were not significantly affected.[1] Evaluation of SCOD data in Table 9-1 and 9-2 indicates that lime pretreatment (pH 10.8 to 11.2) results in from about 25 percent decrease to 25 percent increase in SCOD. Alum or $FeCl_3$ pretreatment consistently resulted in 40 to 50 percent removal of SCOD. It is reasonable to expect that such significantly different pretreatment results would affect subsequent carbon treatment responses. It is the authors' conjecture that alum or ferric chloride treatment removes unadsorbable and/or very difficult to adsorb organic material, whereas, lime treatment does not. Whether or not this conjecture is true, the single-stage data in Figures 9-1 and 9-3 should be shifted to lower equilibrium SCOD values to ascertain any effect of staging.

Biological Effects

A biological organic removal mechanism has been inferred based on data presented in Figure 9-1, where organic removal in the carbon system was seen to be considerably higher than predicted by laboratory equilibrium adsorption isotherm tests.

The existence of a biological organic removal mechanism is also verified by data from Campaigns VL1-3 through 6 (Table 9-1). During Campaigns 3 and 5, no carbon was fed to the carbon contactor and yet about 30 percent removal of SCOD was observed. During these campaigns about 150 to 200 lb of spent carbon slurry, of about 8 g/ℓ concentration, was maintained in the carbon contactors. Campaigns 4 and 6, with carbon dosages of 98 and 31 mg/ℓ, were run parallel to Campaigns 3 and 5, respectively. Using biological organic removal data from Campaign 3 and 5, it is inferred that 66 and 82 percent of the total SCOD removal achieved in Campaigns 4 and 6, respectively, was by a biological mechanism.

Unfortunately, there was no parallel operation with significant difference in SRT which would have allowed making an absolute identification of SRT effects. However, by comparing Campaigns VL1-1 and 2 with Campaigns VL1-7' and 8, an effect of SRT can be seen for single-stage operation. Going from VL1-1 and 2, to VL1-7' and 8 it is evident that equivalent effluent quality was achieved at a carbon dosage of 141 mg/ℓ and an SRT of 7.1 days as was achieved at a carbon dosage of 309 mg/ℓ and an SRT of 2.8 days.

Digressing for a moment, it is germane to note that when all pertinent treatment variables are held nearly constant, the carbon system response was precisely reproduced. Comparison of SCOD removal for parallel Campaigns VL1-1 with VL1-2 and VL1-7' with VL1-8 demonstrates this precision.

By comparing Campaign VF2-12 with Campaign VF2-13 an effect of SRT can also be seen for two-stage operation. Going from VF2-12 to VF2-13 it is evident that by reducing carbon dosage from 108 mg/ℓ to 75 mg/ℓ and increasing SRT from 9.8 days to 20 days equivalent effluent quality of about 3 mg/ℓ SCOD was achieved.

Regeneration Effects

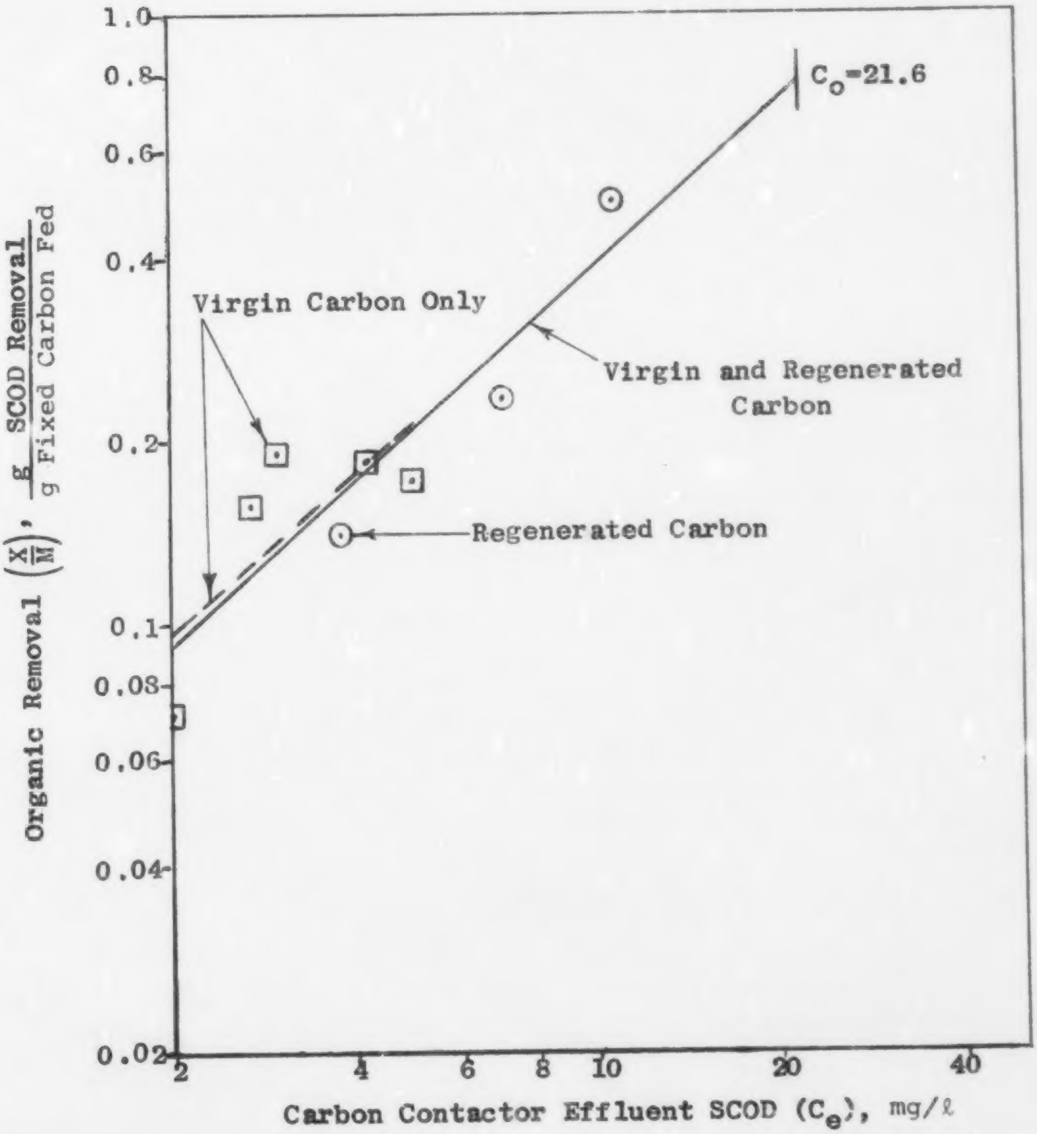
A reduction in SCOD adsorptive capacity of regenerated carbon was shown in Section VIII by iodine number data and by comparative equilibrium adsorption isotherm tests.

Figure 9-5 is a Freundlich plot of all virgin and regenerated carbon data (two-stage counter-current contacting). Regression curves are shown for the virgin carbon data and for virgin plus regenerated carbon data. In the same range of equilibrium SCOD concentration, regenerated carbon organic removals were slightly below virgin carbon organic removals. The limited number of data points and the relatively high variance of the virgin carbon data precludes identification of a statistically significant difference. The net effect of a slight loss of adsorptive capacity is minimized when reusing regenerated carbon in the plant due to the presence of a substantial biological organic removal mechanism.

Comparison with Previous Results

In a previous contract report, the first and second stage organic removal responses were drastically different.[1] In the present study, organic removal models are approximately the same for both stages. During this study high concentrations of carbon fines (50 to 70 mg/ℓ) were allowed to overflow the first-stage contactor-clarifier into the second-stage unit. By so doing, a continual seeding of the second-stage with active biomass was probably achieved. Data for two-stage contacting in Table 9-2 indicate that

FIGURE 9-5: SCOD REMOVAL-ADSORPTION RESPONSE: COMPARISON OF REGENERATED AND VIRGIN CARBON



relatively high SRT's were used. On numerous occasions excessive odor levels were found eminating from the second-stage contactor as well as the first-stage unit. This was not experienced during previous contract studies.[1] In the previous study adsorption and biological removal was inferred to be operative in the first-stage and predominantly adsorptive removal in the second-stage. In the present study adsorption and biological removal was apparently occurring in both stages.

As noted earlier in this Section, pilot plant results from the present study definitely indicated the presence of a chemical pretreatment effect when comparing lime and alum or ferric chloride usage. Such an effect had been conjectured previously, but it was not identified as having a significant effect on the effluent SCOD quality obtainable with the PAC-PCT process.

Results from both the previous and present studies qualitatively indicated that more efficient use of carbon can be made using a two-stage counter-current contacting system compared to a single-stage system. However, the variance of results in both studies precluded defining the difference with reasonable statistical significance. This short-coming is understandable when one considers that staging would effect only the adsorptive removal response. As indicated previously in this Section, adsorption could account for as little as 18 to 34 percent of the total SCOD removal by the PAC-PCT system. To determine the effect of a treatment variable which has an impact on such a small portion of total SCOD removed from a relatively low strength wastewater would require very precise data.

ULTRAVIOLET MONITORING STUDY

Most carbon system response models indicate that carbon dosage requirement to produce a constant effluent SCOD is linearly related to the feed SCOD concentration. If one could instantaneously measure SCOD, either directly or indirectly, carbon dosage could be varied directly with feed SCOD changes. In addition it would be desirable to have real time monitoring of carbon system effluent to alert operating personnel of undesirably high system effluent SCOD. Such an operation should result in production of a very consistent quality effluent.

Unfortunately, instrumentation for direct real-time determination of SCOD on a plant operation scale is not commercially available. Evaluation of available indirect methods indicated that ultraviolet light (UV) absorption might be of utility. A correlation of UV absorbance of 254 nm and TOC had been determined for a variety of treated and untreated waters and wastewaters.[14] Based on this correlation and the desire for instantaneous organic monitoring, a commercially available UV instrument was installed at the pilot plant. The purpose of the UV instrument installation was to demonstrate the usefulness of UV-absorbance as an operational tool to continuously monitor organic concentration. The primary objective was to determine the correlation between UV absorbance at 254 nm and dissolved organics, and to use UV monitoring to operate the carbon treatment system. After two months of data collection, it became evident that a poor correlation was achieved (Figure 9-6). A laboratory study was designed to identify the variables responsible for the poor correlation observed. (Refer to Appendix C).

The basic problem of achieving a general UV absorbance-organic concentration correlation results from the changing relative concentrations of the organic solutes present in the wastewater. In the PAC-PCT process, relative concentration changes result from at least three basic causes: (1) chemical treatment selectivity, (2) activated carbon adsorption selectivity and (3) natural diurnal organic concentration changes.

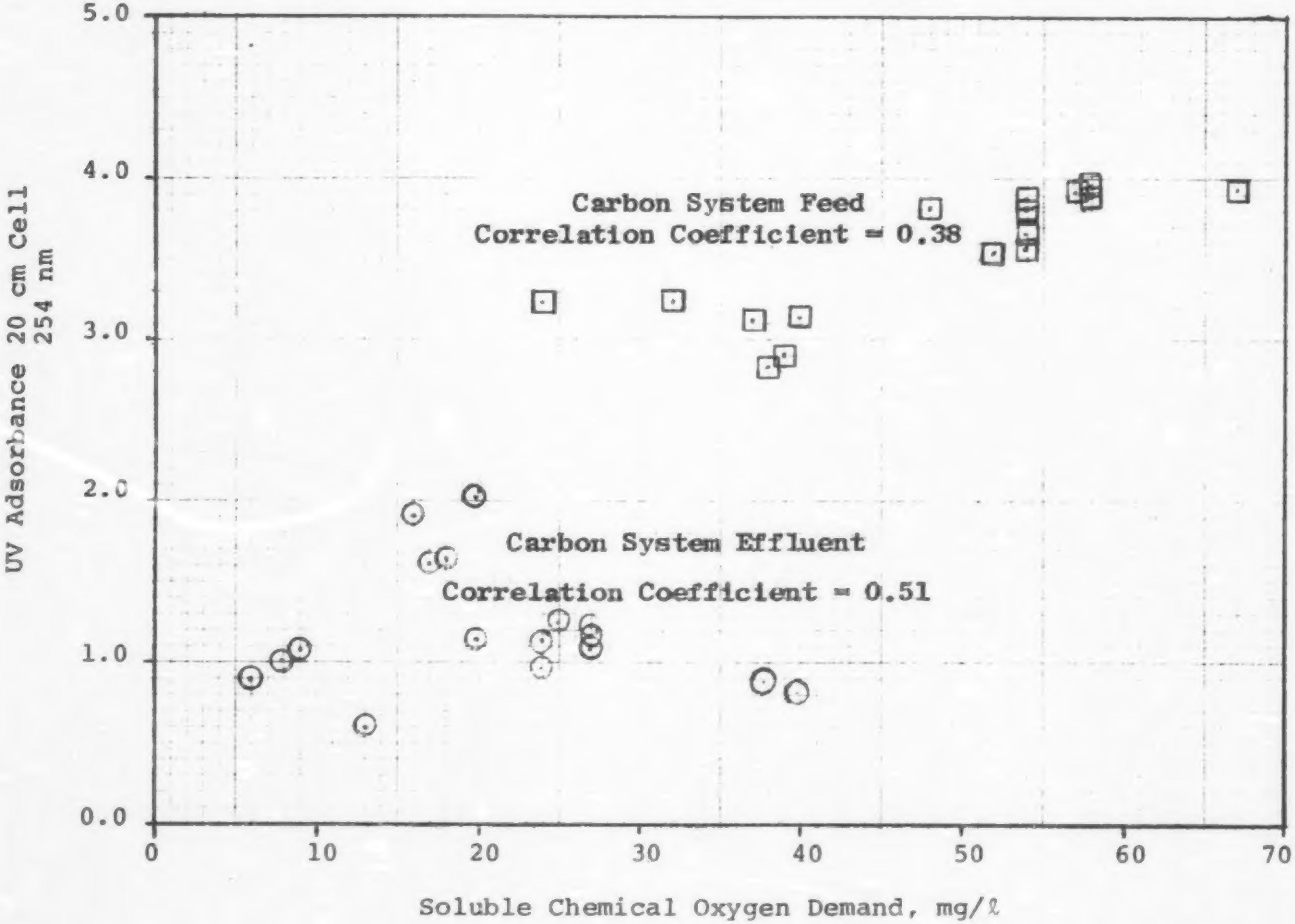
In a selective PAC-PCT treatment process, general UV absorbance-organic concentration correlation prospects are minimal. However, a working correlation should be possible at specific treatment points (e.g., after chemical or carbon treatment) if very consistent treatment effects are achieved.

To identify a working correlation at a specific treatment point, the PAC-PCT plant's final effluent was continuously monitored. Analysis of data for three months of monitoring showed a fairly constant effluent of about 3 mg/ℓ SCOD (1 to 5 range) and 0.4 absorbance (0.2 to 0.8 range). Since the effluent SCOD was such a low and relatively constant value, it was impossible to identify a correlation between UV and SCOD.

It has been generally observed that, after backwash of a granular media filter, poor quality effluent is usually experienced for a short period of time. When this "bad water" was experienced during this study the UV printout increased drastically for a short period of time, indicating UV is responsive to this type of treatment upset.

The use of UV absorbance to monitor the plant effluent for the purpose of determining carbon dosage was judged to be an unrealistic venture.

FIGURE 9-6:
ULTRAVIOLET LIGHT ADSORBANCE VERSUS SCOD CONCENTRATION



SECTION X

PAC-PCT PROCESS DESIGN RECOMMENDATIONS

The purpose of this section is to present the authors' recommendations for design of the various unit operations and processes included in the PAC-PCT municipal wastewater treatment system studied. These recommendations were formulated from the results presented in Sections VI through IX, previous contract work (reference 1) and other relevant experiences of the authors. It is pertinent to note that the following subsections will contain some judgement and opinions of the authors.

The design recommendations presented here will be used in the next section as the basis for an economic analysis of the PAC-PCT system studied.

The unit operations and processes to be discussed will include:

1. Preliminary treatment
2. Chemical treatment
3. Chemical-primary sludge treatment
4. Powdered carbon treatment
5. Carbon regeneration

Also included will be comments concerning operation of the PAC-PCT system.

PRELIMINARY TREATMENT

Preliminary treatment normally consists of screening, grit removal and comminution. During the present study, there was very ineffective preliminary treatment of the raw wastewater being used. Therefore, it was demonstrated that the type of solids-contacting chemical treatment unit used would operate successfully without the removal of either grit or fibrous material. It is, however, recommended that both grit and fibrous material be removed prior to the chemical treatment step to avoid possible clogging and wear of pipes, pumps, and mixing turbines.

Although preaeration of the raw wastewater was not evaluated during this study, it is suggested. It might reduce the strong anaerobic conditions that can exist in the chemical treatment unit, especially when alum or $FeCl_3$ is used.

The following preliminary treatment steps are recommended for Salt Lake City or similar wastewaters.

1. Mechanically cleaned bar screens with 2-inch openings
2. Aerated grit chambers with grit removal and washing
3. Comminution

Conventional design and sizing criteria are suggested. Raw wastewater flow measurement, indication, recording and integration (totalizing) are also considered a must in order to enable pacing subsequent chemical and reagent feed flow rates, thus maintaining constant dosages.

CHEMICAL TREATMENT

Choice of Chemicals

The choice of treatment chemical is often based on engineering judgement and/or preliminary laboratory or pilot plant testing results.

Lime would probably be the chemical of choice for a wastewater low in alkalinity and high in phosphorus concentration. Phosphorus removal by lime treatment is directly related to pH and independent of influent phosphorus concentration. For a low alkalinity wastewater raising the pH can be achieved at reasonably low lime dosages.

If a wastewater is high in alkalinity and low in phosphorus concentration, then either alum or ferric chloride would probably be the chemical of choice. Neither alum nor ferric chloride treatment are adversely affected by high alkalinity except that slightly increased dosages would usually be required due to operation at a pH in excess of the optimum. At a given effluent soluble phosphorus concentration, the amount of either alum or ferric chloride required is directly proportional to the feed phosphorus concentration.

Another alternative would be to use the combination of lime and $FeCl_3$ at a pH range of about 9 to 10. In this approach the lime provides for pH adjustment (phosphorus insolubilization) and the $FeCl_3$ for coagulation of suspended solids (in cluding chemical precipitates).

A desire to remove soluble organics in the chemical treatment unit would also effect the choice of chemicals. During this study, it was observed that use of alum or ferric chloride treatment resulted in 40 to 50 percent SCOD removals; whereas lime treatment resulted in no consistent effect on SCOD removal. The observed SCOD removal values for lime treatment ranged from +25 to -25 percent, indicating net removal at times and net solubilization of particulate matter at other times. The results in Section IX strongly implied that alum or ferric chloride treatment removed some apparently unadsorbable SCOD and thus minimized PAC-PCT system effluent SCOD.

Sludge dewatering and disposal alternatives are key considerations in the choice of chemical. During this study, lime-primary sludge was found to thicken and dewater more easily than either alum or ferric-primary sludges. However, five to six times as much sludge (by weight) was produced by lime treatment than by alum of ferric chloride treatment. For lime treatment of wastewaters containing a relatively high concentration of magnesium, single-stage lime treatment is appropriate. On the other hand, for wastewaters where the magnesium concentration is low, two-stage lime treatment-recarbonation is considered appropriate. Lime is added to the first stage to produce a high pH level in order to precipitate significant magnesium hydroxide for coagulation. The excess calcium added (in the lime) to achieve the high pH is normally precipitated by recarbonation prior to a second stage solids-contacting clarifier. The additional $CaCO_3$ sludge produced in the second stage must be disposed of. Recalcination of the $CaCO_3$ sludge has been found to be economically justified in some applications.

The above comments provide some insight into relevant considerations concerning the choice of treatment chemical. The only rational approach to choosing a treatment chemical is to assess the relative total capital and operational and maintenance costs for both the wet end and the sludge treatment and disposal alternatives available.

A previously reported economic analysis of lime, alum or ferric chloride treatment of Salt Lake City raw municipal wastewater indicated that alum was the least expensive.[1] This analysis included chemical-primary sludge incineration and land disposal of the ash produced. The results of the current study substantiate that alum is the economic choice for Salt Lake City.

It is the authors' judgement that designing for continuous use of polyelectrolyte coagulation-flocculation aid cannot be economically justified. It is, however, readily acknowledged that use of polyelectrolytes allows operation of the chemical treatment clarifier at significantly increased hydraulic loadings. Increased sludge thickening rates would also be expected in both the clarifier and a gravity sludge thickener. It is, therefore, recommended that the chemical treatment clarifier and gravity sludge thickener be designed to routinely operate without the use of polyelectrolyte; but that a polyelectrolyte feeding system be installed. Doing so will afford the plant operator some flexibility and operational alternatives, especially during periods of treatment upsets, unexpected high wastewater flows and/or mechanical problems (e.g., a clarifier being off-stream).

Chemical Mixing

Achievement of effective and economical coagulation requires rapid initial dispersion of coagulant throughout the entire wastewater flow. For both alum and ferric chloride coagulation the flash mixer should be designed to produce a mean temporal velocity gradient, G, of 800 to 1000 sec^{-1}, with a turbine tip speed of about 5 ft/sec. A nominal detention time of 30 seconds at peak flow should be adequate.

For lime addition, it is very desirable to react the slaked lime slurry with recycled sludge prior to contacting the raw wastewater. Adding lime to the recycle sludge prevents gross scaling near the lime feed point. The recycle solids offer a large surface area on which the chemical precipitates can form thus promoting rapid, complete chemical reactions and producing a more stable effluent.

Solids recycle can be achieved either by internal or external pumpage of sludge. The method used during this study was internal recycle. This method is favored because of successful experiences during this study and because it avoids the potential problem of plugging in external sludge pumping and plumbing systems.

When using alum or ferric chloride as primary coagulants, if desired, polyelectrolyte should be dispersed with an inline mixer at a relatively low G value (100 to 200 sec^{-1}). In no case should the polyelectrolyte merely be added to the surface of the flocculation (or reaction) zone since very poor dispersion occurs, lowering effectiveness and increasing the cost of using polyelectrolyte. Manufacturer's recommendations for polyelecltrolyte make-up, storage and handling should be followed.

Type of Equipment

A solids-contact treatment unit, with internal sludge recirculation, is recommended by the authors for lime treatment. This type of solids-contact unit can achieve a high degree of sludge recycle and, therefore, produce a stable water which minimizes scaling problems. For lime treatment, high upflow velocity in the draft tube is required to pick up and recirculate sludge. The sludge rakes can be a standard design normally used in wastewater treatment; however, torque capacity for lime treatment applications should be at least twice that required for alum or ferric chloride treatment applications. The chemical treatment unit must have a surface skimming device to remove accumulated floatables. The skimmed floatables may be combined with the blowdown solids from the chemical treatment unit or concentrated separately in a scum concentrator, depending on the amount of floatables.

For alum and ferric chloride treatment, no benefit was defined, either in the laboratory or full-scale, for solids-contact treatment. However, the solids-contact unit used in this study provided a substantial sludge age which was partially responsible for the presence of anaerobic biological activity in the chemical treatment unit. This biological activity along with chemical coagulation resulted in as much as 40 to 50 percent removal of SCOD.

It is recommended that the classical three step (flash mix, flocculation and sedimentation) chemical treatment approach be used for alum and ferric chloride treatment. This recommendation is based on economics. Since no advantage or disadvantage was quantified for a solids-contact unit, the less expensive three step approach should be used.

A major problem with ferric chloride treatment in the solids-contact unit during this study was due to the existence of anaerobic reducing conditions. Under these conditions, it was found that ferric iron was reduced to soluble ferrous sulfide. This ferrous sulfide reported to the ferric-primary sludge making vacuum filter dewatering virtually impossible. Some soluble ferrous iron overflowed the chemical treatment unit and was presumably subsequently oxidized and post-precipitated as ferric hydroxide. Reduction of ferric to ferrous iron results in inefficient use of ferric chloride. Pre-aeration of the raw wastewater or oxygen addition to the solids-contact unit might eliminate the reduction of ferric iron to ferrous iron.

An important factor in the operation of the chemical treatment unit relates to sludge discharge or blowdown capacity. During the operation of any type of clarification device, there will be times when the sludge level will rise. In order to quickly reduce the sludge level, a sludge blowdown capacity must exist which will allow rapid removal of excess sludge to a surge tank. A surge tank is desirable, especially to reduce hydraulic upset of a gravity sludge thickener. It is recommended that the surge tank have a surge capacity equal to a volume of a three foot depth in the clarifier clarification zone. The ability to transfer this volume in no more than one hour is recommended.

Process Monitoring

Effective, efficient and reliable chemical treatment of wastewater requires conscientious and consistent operator attention. Suitable tools and aids are required to provide the operator with knowledge of system performance and input conditions at any given time. In addition, collection, recording and filing of these conditions will provide a historical data base which is necessary for new operator training and engineering evaluation of the system.

Grab and flow-proportional composite sampling facilities should be provided to enable sampling of chemical treatment unit feed, effluent and underflow streams. Grab sample facilities after flash mixing and flocculation are also desirable. Composite samples can be subjected to analysis for specific pollutants to assess average system performance over an operating period (normally one or two days). Grab samples are required to enable assessment of instantaneous performance or system conditions.

As a minimum, continuous monitoring of chemical treatment clarifier effluent for clarity is suggested. The best practical approach is to use a continuous flow through turbidimeter with indication and recording of effluent turbidity. Recording of effluent turbidity allows assessment of trends in addition to the absolute level.

The primary treatment variable in a lime treatment system is pH. It is recommended that continuous monitoring and recording of pH shortly after mixing and reaction of lime (prior to clarification) be used. An electronic signal from the pH indicator can be integrated with one from the raw wastewater flow meter and the result used to automatically pace lime feed to flow and a desired treatment pH. Difficulties were encountered in this study using such an automated system. These difficulties were considered to be due to the specific instrumentation used.

Flow and slurry concentration monitoring, indication and recording of chemical treatment clarifier underflow should be used. Such information will greatly assist the operator in programming and controlling sludge withdrawal to maximize concentration and thus minimize volume. Magnetic flow meters and sonic sludge density devices are suggested.

Suitable facilities for routinely monitoring chemical inventories and instantaneous feed rates are considered very desirable.

Recommended Design

Table 10-1 is a tabulation of the authors' recommended design for cost effective alum treatment for Salt Lake City or similar wastewaters.

CHEMICAL-PRIMARY SLUDGE TREATMENT

Evaluation of chemical-primary sludge treatment and disposal alternatives and strategies was not within the scope of this study. Thus only those sludge treatment alternatives studied will be discussed at any length.

Sludge disposal methods such as lagooning or landfill should be considered for locations where land area is available, and climate and other environmental conditions are favorable. If local conditions preclude landfilling of concentrated sludges, incineration is the obvious alternative. Both landfill and incineration costs will be presented in Section XI to show economic effects. Recalcination of lime sludge is a possibility, but at present is not considered economically feasible for Salt Lake City or similar wastewaters.

With the possible exception of land spreading of sludges, all other alternatives are very cost sensitive to sludge concentration. In general, the more concentrated the sludge the lower the total sludge treatment and disposal costs. Gravity thickening and vacuum filtration of chemical-primary sludges were the only two methods evaluated to determine, in part, the efficacy of concentrating the sludges studied.

Gravity Thickening

Operation and design of the chemical treatment unit has a pronounced effect on gravity thickening performance. It was shown in Section VII that, in general, the greater the chemical treatment unit blowdown sludge concentration, the greater would be the thickener underflow sludge concentration.

TABLE 10-1

CHEMICAL TREATMENT[a]: RECOMMENDED DESIGN PARAMETERS

Chemical:	
Type	Alum
Alum Dosage, mg/ℓ as Aluminum	13
Polyelectrolyte[b]:	
Type	Slightly Anionic
Dosage, mg/ℓ	0.25
Flash Mix:	
Type	Vertical Turbine
Retention Time, min (at peak flow)	0.5
Mean Temporal Velocity Gradient, G, sec^{-1}	800-1000
Clarifier:	
Type	Flocculating Clarifier
Hydraulic Loading, gpm/sq ft (peak)	0.4
Weir Loading, gpd/ft	10,000
Flocculation Detention Time, min. (at peak flow)	15
Flocculating Mean Temporal Velocity Gradient, G, sec^{-1}	20-60

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[a] Of Salt Lake City or similar wastewater

[b] Routine use not recommended

The chemical treatment unit blowdown sludge concentration was effected, at least, by diurnal flow variations and by use of polyelectrolyte.

To minimize thickener feed flow variation, it is recommended that intermittent sludge blowdown from the clarifier be to a surge tank. The gravity thickener would then be fed from the surge tank by a feed pump at a near constant rate.

It is recommended that the gravity thickener be designed to achieve maximum possible underflow concentration. However, excessively long SRT's should be avoided. SRT's in excess of about 3 days may result in septic conditions for alum or $FeCl_3$-primary sludges and could adversely effect thickener performance and subsequent dewatering operations.

It is recommended that polyelectrolyte feed to the gravity thickener be designed for. Polymer would be used only when needed to maintain a maximum feed concentration prior to vacuum filtration (if used).

Pickets on the thickener rake arms are recommended. Pickets provide channels for water released from the thickening sludge.

Vacuum Filtration

In general, vacuum filtration yield is directly proportional to the feed sludge concentration. Also, cake moisture content will decrease as feed sludge concentration increases. Thus, it was recommended in the previous subsection to design the gravity thickener to produce the maximum possible underflow concentration.

Vacuum filtration dewatering of lime-primary sludge was found to be easy and effective without the use of chemical conditioning. The use of polyelectrolyte for conditioning dilute lime-primary sludges was observed to substantially increase filtration yield. It is, therefore, recommended that polyelectrolyte feeding facilities be provided for lime-primary vacuum filtration systems to provide some operational and performance flexibilities.

About 30 to 40 percent lime $\{Ca(OH)_2\}$ by weight was required to effectively dewater alum-primary sludge. Adequate vat mixing of lime conditioned alum-primary sludge is critical to prevent solids settling out.

As indicated previously, the presence of FeS precipitate in the ferric-primary sludge made vacuum filtration dewatering of ferric-primary sludge nearly impossible. If FeS formation can be avoided then vacuum filtration dewatering, with lime conditioner, should be possible.

Recommended Design

In the design of any sludge thickening and vacuum filtration dewatering system, operational strategies for handling various types of system upsets must be considered. The desirability of surge capacity ahead of the gravity thickener has already been discussed.

The vacuum filtration station should be designed for excess capacity either by planning for less than continuous operation or by provision of stand-by units. Ample spare parts, including filter media, should be provided.

In general, the authors recommend designing for filtration rates significantly lower than average achievable rates. The filter station must be able to process all sludge produced, in a reasonable time frame and for conditions of poorest predictable dewatering properties.

Table 10-2 is a tabulation of recommended design criteria for gravity thickening and vacuum filtration of alum-primary sludge for Salt Lake City or a similar wastewater.

TABLE 10-2

ALUM-PRIMARY SLUDGE:
DESIGN RECOMMENDATION FOR SLUDGE DISPOSAL

Gravity Thickener	
Raking	constant speed with pickets
Solids Loadings, lb/day/sq ft	10
Predicted Underflow Solids, g/ℓ	50
Vacuum Filter	
Type	standard belt filter with flocculator
Media Type	Nylon (NY 317-F)[c]
Maximum Lime Feeding Capacity, % by wgt	50
Yield, lb/hr/sq ft[a]	2
Cake Moisture Content, % by wgt[b]	75

[a] Based on alum and primary sludge solids only

[b] Based on total cake solids

[c] Eimco BSP Division of Envirotech Designation

CARBON TREATMENT

Carbon Contactor

One of the most important design requirements for a carbon treatment system is to establish the PAC dosage necessary to achieve a given effluent quality. Since adsorption, biological and coagulation removals all occurred in the PAC-PCT process studied, there is no obvious way to establish a carbon dosage. The reponse of other wastewaters to PAC-PCT treatment can only be conjectured. Therefore, all that will be done in the next section is to show the relative economics of using various carbon dosages.

Since pilot plant results indicated that as little as 26 percent of SCOD removed across the entire plant was removed by adsorption, it would appear that a lower grade, and/or lower cost carbon than used in this study could be used. About 43 percent of the removal was by a biological mechanism in the carbon contactors.

All of the results reported thus far for these pilot plant studies were for a relatively weak municipal wastewater. The response that may be obtained from other stronger wastewaters might be quite different than those observed during this study. It should be emphasized that long term field data (pilot plant and/or full scale) is necessary to identify all of the soluble organic removal mechanisms operative in a PAC-PCT system.

The effect of counter-current staging is still not well defined. It was suggested in a previous report that two-stage counter-current contacting enhanced stable plant operation.[1] The second stage adds liquid-solids separation reliability (at least redundancy) and longer SRT's. Two stages are more costly and, of course, involve twice the equipment to operate and maintain. When a low level of adsorptive removal is sufficient, a single-stage contactor would be the less expensive approach.

If the major SCOD removal mechanism were adsorption, it is probable that two-stage counter-current contacting of the adsorbent would be the most economical approach.

The solids-contact treatment units used in this study were judged to be excellent powdered carbon contactors. High concentrations and large volumes of carbon slurries were kept in contact with the wastewater. The units acted, in part, as suspended growth anaerobic bio-reactors.

Because of the very small particle size (95 percent smaller than 325 U.S. Mesh) and density (1.52 g/cc) of powdered carbon, requirements for low hydraulic loadings would be expected. Powdered carbon has been shown to be self flocculating at slurry concentrations above 1000 mg/ℓ. Peak overflow rates of 0.7 - 0.9 gpm/sq ft appear to be reasonable based on the results of this study.

Once through flash mixing, flocculation (contact time) and sedimentation systems have been used for laboratory and pilot plant scale PAC studies. These systems did not include PAC recycle and therefore did not experience significant biological activity and thus removal of some SCOD by this mechanism. Usually considerable amounts of coagulating/flocculating chemical additives are required for the once through approach.

The authors consider the use of solids-contacting type of units for PAC contacting/clarification as the method of choice.

Biological removal of SCOD, in the solids-contact units, was accomplished predicated by maintaining powdered carbon SRT's in the order of several days. Maintaining SRT value in excess of more than 3 days during warm summer months and about 7 days during cold winter months was undesirable due to the production of offensive gaseous odors such as H_2S. The need for exercising control of SRT may require deviation from use of standard solids-contactor unit sizes. Specifically the volume of the reaction zone (i.e., size of mechanism) would be based on the desirable carbon SRT for a given application. It would, of course, not be considered desirable to reduce the nominal carbon-influent wastewater flow contact time to less than about 10 to 15 minutes. Otherwise adsorptive equilibriums would not be closely approached.

There were numerous full scale engineering design considerations not evaluated in the current study. For example, requirements for virgin carbon make-up, handling and feeding systems, materials of construction, mixers, pumps, etc., were not systematically evaluated. It is suggested that carbon make-up, handling and feeding technology implemented by large potable water treatment operations using PAC for taste and odor control may be directly applicable to wastewater applications. The same suggestion may be valid for mixers, pumps, etc., also.

It can be noted that after some two years of use the mild steel solids-contact treatment units showed no obvious signs of erosion and/or corrosion. The surfaces of these units were originally prepared by sand blasting and painting with one coat of a phenolic fortified alkyd primer and then two coats of coal tar epoxy paint, eight mils dry thickness each.

Granular Media Filtration

Filtration is the final liquid-solids separation step in the PAC-PCT system studied. Design for achievment of no more than 5 mg/ℓ average suspended solids is recommended. This will allow minimizing loss of expensive carbon and assure a clean, esthetically pleasing system effluent.

To produce a low suspended solids effluent consistently and reliably, conservative hydraulic loading is recommended. Peak filtration rates should be no more than 3 to 4 gpm/sq ft.

The suspended solids in the filter feed were observed to produce a "weak floc" when removed by filtration. Therefore sudden and/or large changes in filtration rate should be avoided to prevent breakthrough of suspended solids into the filter effluent. In addition the limitation of terminal headloss to no more than 8 ft of H_2O would be recommended.

Use of anionic polyelectrolyte, added by in-line mixing just ahead of the filter inlet, was found to result in a slightly "tougher" floc which was less susceptible to premature breakthrough. It is, therefore, recommended that provisions be made for feeding polyelectrolyte when rapidly varying or high flow rate conditions are anticipated.

Considerable experience in the area of designing and applying granular media filters in wastewater applications have convinced the authors of the efficacy of using auxiliary air-scour, prior to water backwashing. Achievement of effective backwashing is a prerequisite to any successful filter operation and performance especially in wastewater applications.

In a previous report it was recommended that filter cycle times in excess of about 24 hours should be avoided.[1] The purpose was to minimize potential "mud ball" formations which might not be destroyed by normal backwashing procedures. This recommendation is still valid.

Backwash water should be collected in a mixed surge tank and then slowly recycled back to the first stage of the carbon contacting system. Recycling to the carbon system is necessary to prevent losses of expensive carbon, but can cause a build-up of carbon and inert fines. Inert fines (inorganic) would be produced by carbon regeneration operations. These carbon and inert fines must eventually be bled out of the closed circuit carbon treatment system. A suggested approach is to periodically recycle backwash water to the chemical treatment clarifier feed stream.

Recommendations for filter bed media and backwash procedures will be presented in a subsequent subsection.

Process Monitoring

The solids-contact units recommended for PAC contacting-clarification require the same operational attention and approach as previously suggested for the chemical treatment unit.

The suggestion for continuous real time monitoring of carbon contactor effluent clarity needs to be qualified. The use of turbidimetric devices utilizing white light scattering and/or transmittance principals of operation are not applicable to PAC slurry concentrations of more than about 20 to 30 mg/ℓ suspended solids. Of the two alternate principals of operation the light transmission devices are preferred. Even then some sort of dilution of the clarifier effluent sample stream with suspended solids free water may be required in order to provide a reasonable degree of precision.

The efficacy of the UV monitoring of carbon system feed and effluent for determination of soluble organics was discussed in Section IX. As indicated there, the use of UV monitoring was not considered as a useful operational tool for the PAC system studied.

Continuous sensing and recording of carbon system effluent UV absorption can provide a historical system performance record. An indication of trends in increasing or decreasing UV absorption could be useful operational intelligence in assessing daily or weekly variations in carbon system performance.

The most important operational strategy concerning the carbon system relates to maintenance of some level of biological activity; but not a level which will result in production of undesirable odors.

If lime is used in the chemical treatment step, close monitoring and control of carbon system feed pH is desirable. Wide variations in pH can significantly reduce biological activity for several hours to several days, depending on the severity and duration of a pH variation.

Maintenance of a well mixed (i.e., uniform suspension) carbon slurry in the carbon contactors is considered desirable, even though it may result in relatively dilute underflow concentrations. Maintaining a well mixed slurry is considered conducive to enhancement of biological activity.

The basic approach recommended for controlling biological activity in the carbon contactors is the same as used in the rational operation and control of the activated sludge process, i.e., control of SRT. The basic assumption required is that the active bio-mass present is directly related to the total suspended solids (the same as for activated sludge systems).

Control of SRT necessitates routine knowledge of the inventory of solids in the carbon system and the flux of carbon through the system. The flux of carbon (lb/day) can be determined from the rate of removal of carbon solids in the contactor-clarifier underflow or the feed rate of carbon to the system. The inventory of carbon (lbs) is determined by suspended solids profiles and slurry volume measurements. The carbon inventory value (lbs) divided by the carbon flux rate (lb/day) indicates the carbon system SRT.

During this study operating charts were posted relating carbon system inventory to reaction zone suspended solids and carbon slurry depth (i.e., sample tap). Also posted was a chart relating carbon flux rate to feed carbon dosage and system flow rate. Use of these charts and routinely collected operational data allowed essentially instantaneous determination of an approximate carbon system SRT.

It is anticipated that the granular media filter station will be designed for automatic backwashing. Initiation of backwashing will be based on one of the following criteria:

1. Effluent turbidity exceeds a predetermined level.
2. The filter pressure drop reaches the maximum available.
3. The length of filter run reaches a predetermined value.
4. Operating personnel desire to manually initiate a backwash cycle.

The first three criteria can be automatically evaluated by process evaluation instrumentation. Effluent clarity can effectively be determined by continuous flow through process turbidimeters. The measured turbidity should be indicated and recorded. The pressure drop across the filter media can also be automatically measured by differential pressure cells and the resulting measurement should also be indicated and recorded.

It is also recommended that facilities be installed to, when desired, determine pressure profiles (i.e., pressure versus bed depth) of clogged filters. If excessive penetration of solids into the filter bed is observed, the operator can be alerted to the need to start feeding or increase the feed rate of polyelectrolyte.

Frequent monitoring and assessing the effectiveness of backwashing is recommended. Backwash rate may have to be adjusted due to seasonal variations in water temperature to insure achievement of a minimum desirable filter bed expansion which is recommended to be 20 percent. The frequency of backwashing (i.e., cycle time) can be very simply monitored by installation of a counting device which is actuated by a valve operator. Routine monitoring of clean bed headloss (immediately after backwashing) is desirable to determine any long term trends of increasing headloss. This observation would possibly indicate inadequate backwashing or accumulation of coal fines at the surface of the coal layer.

Recommended Design

Table 10-3 presents key design recommendations for a carbon contacting-clarification and granular media filtration system for treating Salt Lake City or similar wastewater.

TABLE 10-3

CARBON TREATMENT:
PROCESS DESIGN RECOMMENDATION

Carbon Contactor-Clarifier	
Type	Solids contact with internal solids recycle
Peak Hydraulic Loading, gpm/sq ft	0.8
Slurry Concentration, g/ℓ	10
Predicted Underflow Concentration, g/ℓ	30
Approximate Carbon SRT, days	3-7
Minimum Reaction Zone Hydraulic Retention Time, Minutes	15
Granular Media Filtration	
Peak Hydraulic Loading, gpm/sq ft	3-4
Average Backwash Recycle, % of Filtrate Flow	3
Backwash Procedure:	
Air-Scour Rate, SCFM/sq ft	3-5
Air-Scour Duration, minutes	2-10

Table 10-3 (continued)

Backwash Procedures (continued):	
Water Volume, gal/sq ft	70-90
Bed Expansion, % of Static Bed Depth	>20
Filter Media, Size Range, mm	
Coal	1.5-2.7
Sand	0.75-0.84
Garnet	0.25-0.50

The filter media sizes and types indicated in Table 10-3 should not be considered absolute. They simply indicate a design found in this study to produce acceptable performance.

CARBON REGENERATION

Consideration of treatment and cost effectiveness of alternative methods of powdered carbon regeneration is not within the scope of this study. Therefore, only recommendations relating to the use of a fluid bed furnace approach will be presented.

Due to limited knowledge of the authors in the area of fluidized bed furnace technology, it would be presumptuous to make any recommendations without qualification. Certain relevant comments and observations are, however, in order.

Thickening and Dewatering

It was shown in Section VII that the vacuum filtration yield (lb of dry solids/hr sq ft) for spent carbon slurries was directly proportional to feed solids concentration. It is therefore recommended that spent carbon slurries be gravity thickened to the maximum possible concentration prior to vacuum filtration dewatering. Though a rigorous cost analysis was not made, the authors conjecture that one would show the above recommendation to be the most cost effective.

During this study the spent carbon gravity thickeners were located in an enclosed building. The units were routinely underloaded and were often used to accumulate and store spent carbon prior to a regeneration run. No significant odor problems were experienced. When thickener underflow was stored in an inventory tank for up to 2 to 3 weeks and then vigorously mixed, considerable sulfide odors were observed. It would be expected that a full scale operation would not necessitate storage of spent carbon for periods of 2 to 3 weeks and thus odors would not be an operational problem.

As noted in Section VII, carbon sludge vacuum filter yields were lower when using alum or ferric chloride pretreatment than when using lime pretreatment. During the current study only limited closed circuit operation of the carbon system was achieved. It is unlikely that a steady state solids and carbon condition was reached. It is therefore suggested that a conservative interpretation of carbon dewatering performance results can be made.

Poor solids capture was often experienced during this study, especially when polyelectrolyte was not used. Therefore, routine use of a cationic polyelectrolyte is therefore recommended in order to maximize solids capture.

It was found that the spent carbon filter cakes cracked excessively during the dry cycle. The vacuum filter used had a single form and dry vacuum receiver resulting in reduced operating vacuum levels. It is strongly recommended that separate form and dry vacuum systems be used to allow maximum vacuum levels for cake formation and maximum filter yields.

During this study transport of filter cake from the vacuum filter to the fluid bed furnace was accomplished by collection in a hopper, reconstitution as a paste with a ribbon screw and transfer in a pipeline via a Moyno type pump. This transport system worked well normally, but on occasion bridging of filter cake in the hopper throat (6 inches wide by 4 feet long) was experienced. Reconstitution of cake to a paste and maintenance of at least 12 ft/min pipe velocity was considered necessary to effectively transfer cake.

Fluid Bed Furnace

Results from this study indicate that high recovery of fixed carbon is possible with the fluidized bed furnace used. An average recovery of just over ninety percent was experienced (range from 76 to 100 percent). Regenerated carbon had slightly reduced equilibrium adsorption capacity for iodine, molasses and waste soluble COD, when compared to virgin carbon.

In general it was concluded that the fluid bed furnace system evaluated was capable of effectively regenerating powdered carbon spent in the PAC-PCT process studied. There is no basis for concluding that regeneration effectiveness was maximized or that regeneration costs were minimized in this study. Additional study is required to accomplish these two objectives.

It was found in this study that positive control of oxygen in the fluid bed was necessary to achieve successful performance. The approach used during this study to monitor oxygen was considered poor and it is recommended that a more highly instrumented and automated approach previously used by others be considered.[2]

Based on the results of this study use of the bed-injection-gas principal of operation cannot be recommended. The potential cost benefit associated with use of this operating principal suggests, however, that additional development studies may be warranted.

In recommending the off-gas-recycle principal of operation for a fluid bed furnace approach to powdered carbon regeneration, it is recommended that the following design objectives be strived for:

1. Maximize fire box operating temperature consistent with available materials.

2. Minimize fluid bed operating temperature consistent with achievement of an acceptable level of regneration effectiveness.

3. Maximize the quality of bed fluidization (e.g., maximum numer of small orifices in tuyeres; maximum economical bed depth to diameter, etc.).

During this study it was found desirable to remove furnace sand and other foreign material from the recovered carbon by cyclone and screening operations. These steps are also recommended for a full-scale system. Doing so will help prevent plugging of pipes and pumps and minimize the build up of inerts within the closed-loop carbon regeneration and reuse system.

Inorganic precipitates which enter the carbon contacting system via the effluent from the chemical treatment clarifier will report to the spent carbon slurry (i.e., regeneration system feed). Most of these materials will be converted to metal oxides in the furnace operation and thereafter remain with the regenerated reused carbon in the same form. Build up of these materials will ultimately necessitate blowdown from the closed-loop system.

Blowdown of accumulated inorganic ash can be accomplished by wasting along with associated carbon solids. An alternative would be to acid treat recovered regenerated carbon and recycle the leachate to the chemical treatment step.

Design Recommendations

Basically the carbon regeneration system should be designed for continuous operation. Recognizing the need for down time for equipment repairs and maintenance it is recommended that about 15 percent excess capacity be designed into the furnace and about 25 percent excess capacity into the vacuum filter.

Based on the results of this study Table 10-4 presents recommended process design criteria for the powdered carbon regeneration system major components.

TABLE 10-4

CARBON REGENERATION SYSTEM: RECOMMENDED PROCESS DESIGN

Gravity Thickener	
Solids Loadings, lb/day/sq ft	20
Predicted Underflow Concentration, g/ℓ	120
Vacuum Filter	
Type	Standard Belt Filter
Drum Submergence, %	33
Media Type	Polypropylene
Vacuum Level*, in. of Hg	20
Airflow Rate, ACFM/sq ft	~5
Polyelectrolyte:	
Type	Cationic
Dosage, lb/TDS	5-10
Yield, lb/hr/sq ft	5
Cake Solids, %	~27
Filtrate Solids, g/ℓ	<12
Fluidized Bed Furnace (Off Gas Recycle)	
Solids Loadings, lb/hr/sq ft	3
Free Board Velocity, ft/sec	1.2
Firebox Temperature, °F	2000
Operating Temperature, °F	1250

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*Use separate form and dry vacuum receivers

SECTION XI

ECONOMIC ANALYSIS

The following economic analysis is presented for the purpose of showing the estimated costs of a PAC-PCT system for treating a municipal wastewater similar to Salt Lake City, Utah. To enable assessment of economy of scale, estimated costs will be presented for average design flows of 5, 10 and 50 mgd. Peak to average design flows of 1.5 to 1.0 were assumed.

The cost estimates are presented for the chemical (i.e., alum) treatment step, the carbon treatment step and then chemical and carbon treatment. Doing so enables identification of the costs for removal of suspended solids and phosphorus by chemical treatment separately from the cost for removal of soluble organics (BOD_5 and SCOD).

Table 11-1 presents the assumed pertinent raw wastewater and system effluent water qualities. As footnoted in Table 11-1 the actual effluent organics to be expected from a PAC-PCT system will depend on specific raw wastewater soluble organics and their amenability to such a treatment process.

TABLE 11-1

ASSUMED AVERAGE RAW WASTEWATER AND EFFLUENT WATER QUALITIES

Parameter, Units	Raw Wastewater	Chemical Treatment Effluent	Plant Effluent
BOD_5, mg/ℓ	125	25	5[a]
Suspended Solids, mg/ℓ	150	25	5
Turbidity, JTU	50	10	3
Total COD, mg/ℓ	175	40	10[a]
Soluble COD, mg/ℓ	45	25	6[a]
Phosphorus, mg/ℓ as P	6	1	0.5

[a] Will depend on specific raw wastewater soluble organics

Major process design parameters and equipment type and sizing were based on those presented and discussed in Section X.

Assumed unit costs for chemicals and operating costs are presented in Table 11-2.

TABLE 11-2

ASSUMED UNIT COST FOR
ECONOMIC ANALYSIS

Chemicals - (as of January, 1974; FOB Salt Lake City)

Alum	3.5¢/lb
Lime	1.0¢/lb
Polyelectrolyte	$1.50/lb
Powdered Carbon	10.0¢/lb

Capital Costs

All equipment costs were estimated as installed equipment by the authors' firm and amortized at 6 percent for 25 years.

Operating Costs - (as of January, 1974)

Power	$4.00/MG
Ash and Sludge Cake Haulage	$5.00/ton
Maintenance	1% of capital costs/year

CHEMICAL TREATMENT COSTS

Equipment design parameters and assumed operating schedules used to size equipment for the chemical treatment step are shown in Table 11-3. It is apparent that at least two (parallel) flocculating clarifiers are used for all plant sizes. For the 5 and 10 mgd plant flow sizes the solids handling systems have only a single train of equipment. It should also be noted that different solids handling system operating schedules were assumed for the three plant flows. The approach invoked here was that larger plants with larger investments in capital equipment and facilities should strive for approaching near continuous solids handling operating to minimize total treatment costs.

Table 11-4 presents estimated alum treatment costs. Pretreatment costs shown include bar screening, aerated grit removal, comminution and flow measurement lumped together. Raw wastewater pumping was not assumed to be required.

TABLE 11-3

DESIGN PARAMETERS AND MAJOR EQUIPMENT SIZING FOR CHEMICAL TREATMENT

Item	PLANT FLOW, mgd		
	5	10	50
Alum Dosage, mg/ℓ {$Al_2(SO_4)_3 \cdot 17H_2O$}	125	125	125
Alum Flash Mixing:			
Retention Time, minutes @ peak flow	½	½	½
Mixer:			
Type	- All Vertical Radial Flow Turbine -		
Horsepower	5	7.5	40
Tank			
Number	1	1	2
Diameter x depth, ft	7 x 9	10 x 10	13 x 13
Flocculating Clarifiers:			
Peak Hydraulic Loadings, gpm/sq ft	0.4	0.4	0.4
Number of Units	2	2	4
Diameter of Units, ft	100	140	215
Flocculator Horsepower	2	3	5
Gravity Sludge Thickener:			
Solids Loading, lb dry solids/day-sq ft	10	10	10
Number of Units	1	1	2
Diameter x depth, ft	30 x 10	45 x 10	68 x 10

Table 11-3 (continued)

Item	PLANT FLOW, mgd 5	10	50
Vacuum Filtration:			
(Belt Filter)			
Lime Dosage, percent by weight	40	40	40
Yield (including chemical), lb dry solids/hr-sq ft	2.8	2.8	2.8
Operating Schedule:			
Days/Week	5	5	7
Hours/Day	8	16	20
Number of Units	1	1	3
Diameter x face width, ft	12 x 18	12 x 18	12 x 16
Sludge Incineration:			
Type of Unit	- - - Multiple Hearth Furniture - - -		
Solids Loading, lb dry solids/hr-sq ft	8.1	8.1	13
Feed Moisture Content, percent by weight	75	75	75
BTU Values, BTU/lb dry solids	3300	3300	3300
Operating Schedule:			
Days/Week	5	5	7
Hours/Day	8	16	20
Number of Units	1	1	1
Diameter of Units, ft	16.75	16.75	22.25

TABLE 11-4

ESTIMATED ALUM TREATMENT COSTS

Capital Costs (1000 of $)	Alum Treatment with Incineration			Alum Treatment without Incineration	
Flow, mgd	5	10	50	5	10
Pretreatment	86	122	385	86	122
Chemical Contact	748	1065	3220	748	1065
Gravity Thickening	67	78	184	67	78
Vacuum Filters	527	549	1368	527	549
Incineration	825	825	1250	-	-
TOTAL	2253	2639	6407	1428	1814
Contingency (50%)	1127	1320	3204	714	907
Total Capital Cost	3380	3959	9611	2142	2721
Amortized Cost, ¢/1000 gal (25 years @ 6%)	14.5	8.5	4.1	9.2	5.8
Operating Costs (¢/1000 gal)					
Treatment Chemicals	3.7	3.7	3.7	3.7	3.7
Sludge Conditioning	0.6	0.6	0.6	0.6	0.6
Incineration (Fuel, Power, Labor)	2.7	2.1	1.3	-	-
Power	0.2	0.2	0.2	0.2	0.2
Maintenance, @ 1% Cap/Year	1.7	1.0	0.4	1.1	0.7
Supervision & Labor	3.0	2.5	0.8	3.0	2.5
Ash or Sludge Cake Haulage	0.2	0.2	0.2	2.2	2.2
Total ¢/1000 gal Operation	12.1	10.3	7.2	10.8	9.9
Total ¢/1000 gal	26.6	18.8	11.3	20.0	15.7

The contigency cost shown in Table 11-4 under "capital costs" is an estimate of costs for engineering, administration and legal in addition to interconnecting piping, valves, instruments and buildings.

The effect of economy of scale and increased operating schedules on capital costs for larger plant sizes is apparent when comparing the estimated costs for 5 and 50 mgd plant sizes. For the 5 mgd plant amortized capital costs are 55 percent of the total costs. For the 50 mgd plant size amortized capital costs are reduced to 37 percent of total costs.

Comparison of total costs for the 10 and 5 mgd plant sizes indicate that, respectively, 20 and 33 percent increases in costs are indicated for including sludge incineration. Analysis of the cost data shown indicates that the vast majority of increased cost for incineration is due to capital costs requirements.

CARBON TREATMENT COSTS

Equipment design parameters and assumed operating schedules used to size equipment for carbon treatment and regeneration are shown in Table 11-5. It should be noted that the equipment sizing (i.e., loading) parameters indicated are maximum desired values. The actual operating schedules indicated results in actual loadings less than the values shown, in some cases.

Table 11-6 presents estimated carbon treatment costs, all with and at 5 and 10 mgd without regeneration. The capital costs shown for the "without regeneration" case could be reduced slightly if spent carbon were wasted to the chemical-primary solids handling system, rather than handled separately as shown in Table 11-6.

Analysis of the estimated costs in Table 11-6 shows that for a carbon dosage of 100 mg/ℓ it is cost effective to use carbon once and throw it away for plant flows less than 5 mgd. For a carbon dosage of 300 mg/ℓ it is cost effective to use carbon once and throw it away for plant sizes less than about 3 mgd.

Because of the equipment intensive carbon regeneration system there is a substantially greater economy of scale for the case where regeneration and reuse is employed than for when it is not.

TABLE 11-5

DESIGN PARAMETERS AND MAJOR EQUIPMENT SIZING FOR POWDERED CARBON TREATMENT

Item	PLANT FLOW, mgd 5		10		50	
Contacting Clarification						
Solids Contact Units:						
Number of Stages	2		2		2	
Peak Hydraulic Loading, gpm/sq ft	0.8		0.8		0.8	
Number of Units	2		2		4	
Diameter of Units, ft	96		135		210	
Granular Media Filters:						
Peak Hydraulic Loading, gpm/sq ft	3		3		3	
Number of Units	2		3		14	
Diameter of Units, ft	34		38		40	
Backwash Collection Tank:						
Diameter x Depth, ft	20 x 12		20 x 14		20 x 16	
Mixer Type	- - - Vertical Radial Flow Turbine - - -					
Mixer Horsepower	2.5		2.5		2.5	
PAC Regeneration						
PAC Dosage	100	300	100	300	100	300
Gravity Thickener:						
Solids Loadings, lb/dry solids/day-sq ft	20	20	20	20	20	20
Number of Units	1	1	1	1	1	1
Diameter, ft	18	28	22	40	50	90
Side Water Depth, ft	8	8	8	8	9	14

Table 11-5 (continued)

Item	100	300	100	300	100	300
Vacuum Filtration: (Belt Filter)						
Cationic Polyelectrolyte Dosage, lb/TDS	5	5	5	5	5	5
Yield, lb dry solids	6	6	6	6	6	6
Operating Schedyle:						
Days/Week	5	5	5	5	7	7
Hours/Day	8	8	20	20	24	22
Number of Units	1	1	1	1	1	2
Size of Units:						
Diameter, ft	6	10	6	8	8	10
Face Width, ft	8	12	6	14	14	16
PAC Thermal Regeneration: (Fluid Bed Furnace)						
Solids Loading, lb dry solids/ hr-sq ft	3	3	3	3	3	3
Feed Moisture, percent by wgt	77	77	77	77	77	77
Bed Temperature, ^{o}F	1500	1500	1500	1500	1500	1500
Fire Box Temperature, ^{o}F	2000	2000	2000	2000	2000	2000
Spacial Velocity, ft/sec	1.2	1.2	1.2	1.2	1.2	1.2
Number of Units	1	2	1	2	2	7
Diameter of Units, ft	20	20	20	20	20	20

TABLE 11-6

ESTIMATED CARBON TREATMENT COSTS

Regeneration and Reuse	YES						NO			
Plant Flow, MGD	5		10		50		5		10	
Carbon Dosage, mg/ℓ	100	300	100	300	100	300	100	300	100	300
Capital Costs (1000 of $)										
Carbon Treatment	668		997		2985		668		997	
Granular Media Filtration	329		580		2916		329		580	
Gravity Thickening	55	74	60	97	115	147	55	74	60	97
Vacuum Filtration	195	330	170	270	270	700	195	330	170	270
Thermal Regeneration	800	1600	800	1600	1600	5600	-	-	-	-
Total	2047	3001	2607	3544	7886	12,384	1247	1401	1807	1944
Contingency (50%)	1023	1501	1304	1772	3943	6174	624	701	904	972
Capital Costs										
Total	3076	4502	3911	5316	11,829	18,558	1881	2102	2711	2916
Amortized Cost[a], (¢/1000 gal)	13.2	19.3	8.4	11.4	5.1	8.0	8.0	9.0	5.8	6.3
Operating Costs (¢/1000 gal)										
Carbon Makeup	0.8[b]	2.5[b]	0.8[b]	2.5[b]	0.8[b]	2.5[b]	8.0[c]	24.0[c]	8.0[c]	24.0[c]
Regeneration (Fuel, Power, Labor)	1.5	3.4	1.5	3.4	0.9	2.4	-	-	-	-
Power	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2
Supervision & Labor	3.0	3.0	2.5	2.5	0.8	0.8	3.0	3.0	2.5	2.5
Maintenance,@ 1%/Cap/Yr	1.8	2.6	1.1	1.5	0.5	0.9	1.1	1.2	0.8	0.9
Total ¢/1000 gal Operation	7.3	11.7	6.1	10.1	3.2	6.8	12.3	28.4	11.5	27.6
Total Treatment Costs (¢/1000 gal)	20.5	31.0	14.5	21.5	8.3	14.8	20.3	37.4	17.3	33.9

[a] 25 years @ 6% [b] 10% Makeup [c] 100% Makeup

TOTAL PAC-PCT TREATMENT COSTS

Table 11-7 and Figure 11-1 show the estimated costs for a PAC-PCT system which includes sludge incineration and powdered carbon thermal regeneration and reuse. As indicated in Table 11-1 such a system is suggested to be capable of producing an effluent quality of comparable BOD_5 and phosphorus concentrations and lower in COD concentration than a conventional biological treatment system followed by chemical treatment for phosphorus removal and granular media filtration for suspended solids removal. As previously noted the efficacy of the PAC-PCT system treatment effectiveness is dependent on the specific raw wastewater soluble organic characteristics (i.e., their amenability to removal by carbon treatment).

Evaluation of the estimated costs for increasing plant sizes in Table 11-7 and Figure 11-1 indicates that similar economy of scale exists for the chemical and carbon treatment systems. This observation is presumably due to the similarity in wet end and sludge and regeneration equipment schemes.

Figure 11-1 shows that above a plant size of about 20 mgd only a nominal economy of scale is realized.

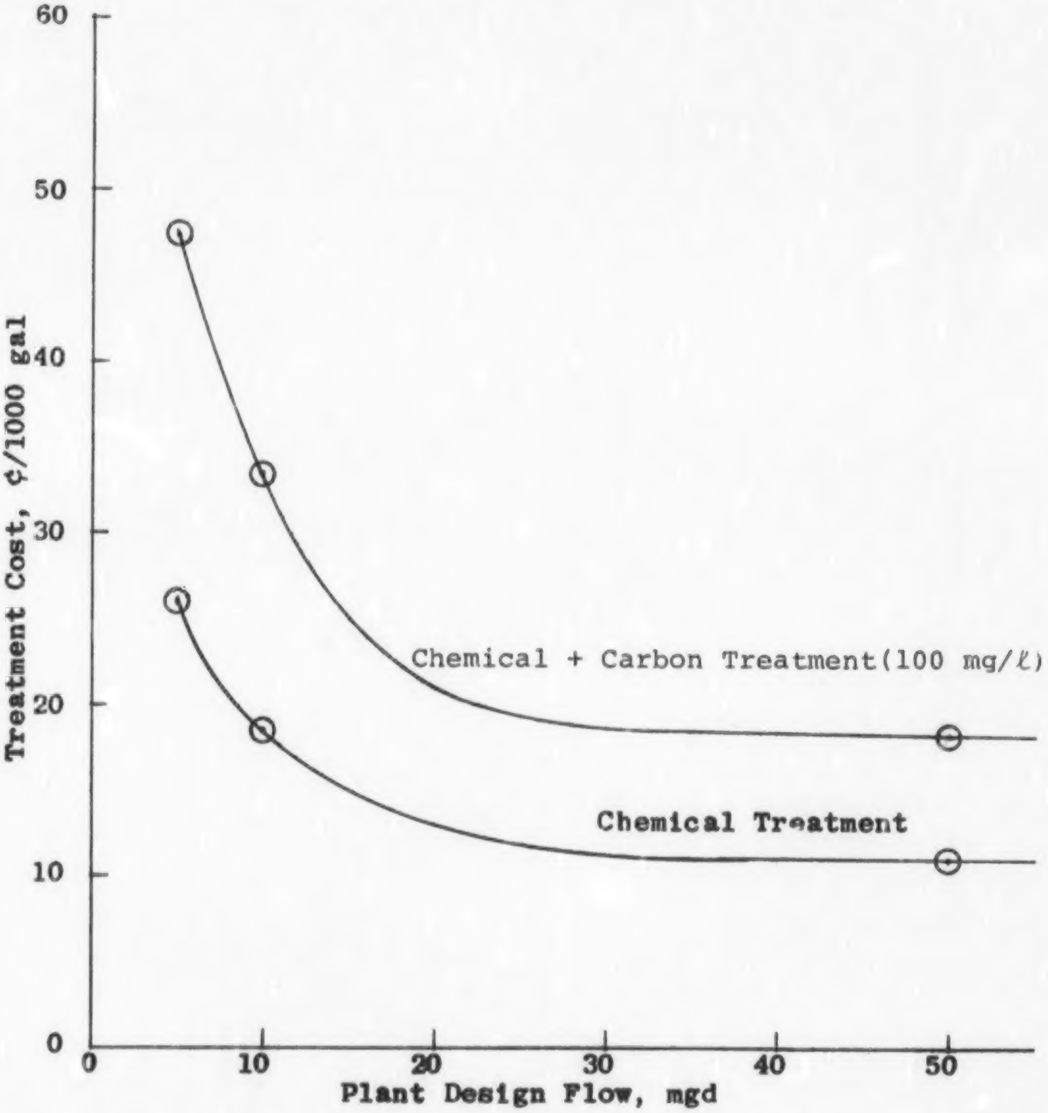
Based on the economic analysis presented it is the authors' judgement that the PAC-PCT process evaluated represents a potentially cost effective alternative for treatment of certain municipal wastewaters.

TABLE 11-7

SUMMARY OF PHYSICAL-CHEMICAL TREATMENT COSTS - WITH ALUM TREATMENT, SLUDGE INCINERATION AND CARBON REGENERATION

Plant Flow, MGD	5		10		50	
Chemical Treatment Costs						
Capital Costs - Amortized 25 years @ 6% (¢/1000 gal)	14.5		8.5		4.1	
Operating Cost (¢/1000 gal)	12.1		10.3		7.2	
Total ¢/1000 gal	26.6		18.3		11.3	
Carbon Dose, mg/ℓ	100	300	100	300	100	300
Carbon Treatment Costs						
Capital Costs - Amortized 25 years @ 6% (¢/1000 gal)	13.2	19.3	8.4	11.4	5.1	8.0
Operating Cost (¢/1000 gal)	7.3	11.7	6.1	10.1	3.2	6.8
Total ¢/1000 gal	20.5	31.0	14.5	21.5	8.3	14.8
Chemical and Carbon Treatment Costs						
Capital Cost - Amortized 25 years @ 6% (¢/1000 gal)	27.7	33.8	16.9	19.9	9.2	12.1
Operating Cost (¢/1000 gal)	19.4	23.8	16.4	20.4	10.4	14.0
Total ¢/1000 gal	47.1	57.6	33.3	40.3	19.6	26.1

FIGURE 11-1: ESTIMATED TREATMENT COSTS VERSUS PLANT SIZE - WITH SLUDGE INCINERATION AND CARBON REGENERATION



SECTION XII

REFERENCES

1. Burns, D.E. and Shell, G.L., "Physical-Chemical Treatment of A Municipal Wastewater Using Powdered Carbon" Environmental Protection Technology Series, EPA-R2-73-264 (August, 1973).

2. Shuckrow, A.J., Dawson, G.W. and Bonner, W.F., "Physical-Chemical Treatment of Combined and Municipal Sewage" Environmental Protection Technology Series, EPA-R2-73-149 (February, 1973).

3. Berg, E.L., Villers, R.V., Masse, A.N. and Winslow, L.A., "Thermal Regeneration of Spent Powdered Carbon Using Fluidized Bed and Transport Reactors" Chemical Engineering Progress, Vol. 67, pp. 154-164, (1970).

4. Shuckrow, A.J., Bonner, W.F., Presecan, W.L. and Kazmierczak, E.J., "A Pilot Study of Physical-Chemical Treatment of the Raw Wastewater at the Westerly Plant in Cleveland, Ohio" presented at the International Association on Water Pollution Research Workshop, Vienna, Austria (September, 1971).

5. O'Brien and Gere Engineers, Inc., "Town of Clay Chemical-Physical Investigation" private communication, Syracuse, New York (January, 1971).

6. Culp, R.L. and Culp, G.L., Advanced Wastewater Treatment, Van Nostrand Reinhold Company, New York, New York (1971).

7. Burns, D.E.., Inter-Company Communication, March 6, 1972.

8. Talmage, W.P. and Fitch, E.B., "Determining Thickener Unit Area" Industrial and Engineering Chemistry, Vol. 47, No. 1, pp. 38-41 (January, 1955).

9. Knopp, P.V., and Gitchel, W.B., "Wastewater Treatment with Powdered Activated Carbon Regeneration by Wet Air Oxidation" presented at the 25th Purdue Industrial Waste Conference (May, 1970).

10. Perry, R.H., Chemical Engineer's Handbook, McGraw-Hill, Inc., (1963).

11. Dayio, K., and Lenenspiel, O., Fluidization Engineering, John Wiley and Sons, Inc., New York (1969).

12. Beebe, R.L., "Activated Carbon Treatment of Raw Sewage in Solids-Contact Clarifiers" Environmental Protection Technology Series, EPA-R2-73-183 (March, 1973).

13. Davis, D.S., and Kaplan, R.A., "Removal of Refractory Organics from Wastewater with Powdered Activated Carbon," Journal WPCF, Vol. 35, No. 3, p. 442 (1966).

14. Dobbs, R.A., Wise, R.H., and Dean, R.B., "The Use of Ultraviolet Absorbance for Monitoring the Total Organic Carbon Content of Water and Wastewater," U.S. Department of the Interior, FWPCA Advanced Waste Treatment Research Laboratory Report, Water Research, Vol. 6, p. 1173 (1972).

15. Standard Methods, 13th Edition, American Public Health Association Inc., Washington, D.C. (1971).

16. ASTM Standards, Part 9, American Society for Testing and Materials, Philadelphia, Pa. (1964).

17. Kynch, G.J., "A Theory of Sedimentation," Trans. Faraday Soc. 48, 166, (1952).

SECTION XIII

GLOSSARY

Abbreviations:

ABS	alkyl benzene sulfonate
BIG	bed injection gas
BOD_5	five-day, 20°C-biochemical oxygen demand
cc	cubic centimeter
COD	chemical oxygen demand
CT	GMF cycle time
cu ft	cubic feet
DC	direct current
EPA	Environmental Protection Agency
FBF	fluidized bed furnace
FFR	form filtration rate
FT	vacuum filter cake form time
ft	feet
fpm	feet per minute
fps	feet per second
GAC	granular activated carbon
gal	gallon
G	mean temporal velocity gradient
GMF	granular media filter
gpd	gallons per day
gpm	gallons per minute
Hp	horsepower
hr	hour
in.	inch
JTU	Jackson turbidity units
kg	kilogram
ℓ	liter
lb	pound
log	logarithm
M	meter
M_c	cake moisture content
mgd	million gallons per day
mg	milligram
MG	million gallons

MHF	multiple hearth furnace
min	minute
ml	milliliter
mm	millimeter
N	normality
nm	nanometer
OGR	off gas recycle
ΔP	incremental pressure
PAC	powdered activated carbon
PCT	Physical-Chemical Treatment
pH	negative logarithm of the hydrogen ion concentration
PM	post meridian
PVC	polyvinyl chloride
P/I	pressure indicator
rpm	revolutions per minute
RZ	reaction zone
SCOD	soluble chemical oxygen demand
SCFM	standard cubic feet per minute (@ 20°C, sea level)
sec	second
sq	square
SRT	solids retention time
SS	suspended solids or single-stage
STOC	soluble total organic carbon
SWD	side water depth
TOC	total organic carbon
TSL	thickener solids loading
UV	ultraviolet light
W	vacuum filter dry cake weight
wgt	weight
Y	vacuum filter yield
2SCC	two-stage counter-current
K	constant in Equation (1)

Symbols:

@	at
°C	degrees centigrade
C_r	raw wastewater solbule organic concentration
C_o	feed soluble organic concentration
C_i	intermediate stage soluble organic concentration
C_e	effluent soluble organic concentration
°F	degrees fahrenheit
M	powdered carbon concentration
#	number
X	organics removed (C_o-C_e)
θ_D	vacuum filter cake dry time
¢	cents
$	dollars
∞	infinite
$\propto$	proportional to
$\lesssim$	greater than or equal to
>	less than
~	approximately
%	percent

English Metric Conversion Table:

Form Filtration Rate (FFR), lb/hr/sq ft = 4.89 kg/hr/sq M

feet (ft) = 0.3048 M

gallon (gal) = 3.785 ℓ

gallons per minute (gal/min) = 3.785 ℓ/min

Horsepower (Hp) = 1.34 kilowatts

hydraulic loadings (gpm/sq ft) = 0.679 ℓ/sec/sq M

inch (in.) = 2.54 cm

inches per minute (in/min) = 2.54 cm/min

million gallons (MG) = 3785 metric tons or M^3

million gallons per day (mgd) = 3785 metric tons/day on M^3/day

pound (lbs) = 454 gms or 0.454 kg

pound per hour (lb/hr) = 0.454 kg/hr

pounds per ton of dry solids (lbs/tds) = 0.05% or 500 ppm

simplified correlating factor (min·sq ft/lb) = 0.205 min·sq M/kg

standard cubic feet per minute (SCFM) = 0.0283 SCM/min

thickener solids loading (lbs/day/sq ft) = 4.89 kg/day/sq M

yield (lbs/hr/sq ft) = 4.89 kg/hr/sq M

SECTION XIV

APPENDICES

APPENDIX A

SAMPLING AND LABORATORY TEST PROCEDURES

SAMPLING

Collection of samples for analytical testing was achieved using one of the following methods:

1. Automatic composite sampling,
2. Routine grab-composite sampling,
3. Non-routine grab sampling, and
4. Reagent feed grab sampling.

Flow proportioned composite samples of each process stream were collected and transferred to a refrigerator using an automatic sampler. These samples were composited over a 24 hour period. Two 24 hour composites were usually combined in the laboratory for analysis as 48 hour composites.

Composite samples of the solids-contact units reaction zone and underflow slurries and thickener overflows were collected by manually obtaining 200 mℓ grab samples on a time proportionate basis (normally each 4 hours). These samples were composited for a 24 hour period.

Concentrations of feed chemicals, spent and regenerated carbon slurries, and sludges were monitored by taking grab samples of well mixed inventoried volumes and analyzing for suspended solids or specific chemicals. Grab samples were obtained to determine instantaneous concentrations of total or suspended solids at various depths in thickeners and solids-contact units. Grab samples of process streams were also used to determine sulfide, iron and aluminum concentrations.

ANALYTICAL PROCEDURES

In certain of the following subsections the term "soluble" is used. By arbitrary definition this term refers to that portion of a given pollutant or material contained in an aqueous sample which will physically pass through the openings in a 0.45 micron membrane filter.

Turbidity

Turbidity of grab and composite samples were determined with a Hach Chemical Company, Model 2100 laboratory turbidimeter. This device is a nephelometer (90 degree light scattering) which was standardized with a plastic rod. Turbidity was reported as Jackson Turbidity Units (JTU).

Suspended Solids

Suspended solids were defined as material retained on a 0.45 micron membrane filter. Concentrated samples of chemical-sewage sludge and carbon slurries were filtered through glass mat filter pads (Type GF/B). A calibrated 30 mℓ syringe was used to measure sludge sample volumes. The Standard Methods procedure for Nonfilterable Residue was used.[15]

Phosphorus

Total phosphorus was determined by the persulfate digestion method.[15] Orthophosphate was determined by the ascorbic acid method.[15]

Chemical and Biochemical Oxygen Demand

Total and soluble chemical oxygen demand (COD, SCOD) and 5-day, 20°C biochemical oxygen demand were determined according to Standard Methods.[15] Total COD was not determined on carbon contactor effluent samples since one mg/ℓ of carbon was found to exert about one mg/ℓ of COD.

pH

The pH data reported were obtained with a laboratory pH meter. Process pH meters and recorders were used only to automatically monitor and control lime feed and acid neutralization during lime treatment.

Hardness, Alkalinity, Sulfides

These analysis were performed according to Standard Methods.[15]

Iron, Aluminum, Calcium, Magnesium

These analysis were made by Atomic Adsorption.

Available Lime Index

"The available lime index" of hydrated lime was determined using the procedure presented in "Standard Analysis of Limestone, Quicklime, and Hydrated Lime".[17]

Gaseous Oxygen, Carbon Dioxide and Carbon Monoxide

Oxygen was determined with a Bacharach 0-7 percent by volume oxygen Indicator, Model CPD, using Fyrite O_2 indicator solution.

Carbon dioxide was determined with a Bacharach 0-20 percent by volume CO_2 Indicator, Model CND, using Fyrite CO_2 indicator solution.

Carbon monoxide concentration as percent by volume was determined with a Bacharach Universal Gas Sampler, Model 19-7016, using CO indicating tubes no. 19-0195.

Total Suspended Solids, Ash, and Volatile Solids

Total suspended solids of carbon slurry samples were determined according to Standard Methods. Ash content of carbon solids was determined according to the procedure presented in ASTM, Part 28 (1964), D1506-59, p. 750. Volatile solids of carbon samples were determined according to the procedure presented in ASTM, Part 28 (1964), D1620-60, p. 811, with the exception that a temperature of 550°C was used, rather than 950°C, to account for organic volatiles only.[16]

Volatile and ash content of chemical sludges were determined according to Standard Methods.[15]

Iodine Number

The iodine number was defined as the milligrams of iodine adsorbed by one gram of fixed carbon when the iodine concentration of the residual filtrate was 0.02 normal.

Procedure:

1. Pulverize an air dried sample of carbon in a mortar and pestle.

2. Dry for a minimum of 3 hours at 105°C.

3. Weigh enough dried pulverized carbon to result in the removal of iodine from 0.1 Normal to about 0.02 Normal.

4. Transfer the weighed sample into a dry, glass-stoppered, 250 mℓ, Erlenmeyer flask.

5. To the flask add 10 mℓ of five percent by weight HCl acid and swirl until carbon is wetted.

6. Place flask on hotplate, bring contents to boil and allow to boil for only 30 seconds.

7. After allowing flask and contents to cool to room temperature add 100 mℓ of standardized 0.1 Normal iodine solution to the flask.

8. Immediately stopper flask and shake contents vigorously for 30 seconds.

9. Immediately filter by gravity through a folded filter paper.

10. Discard the first 20 or 30 mℓ of filtrate and collect the remainder in a clean beaker. Do not wash the residue on the filter paper.

11. Pipette 50 mℓ of the filtrate into a 250 mℓ Erlenmeyer flask.

12. Titrate with standardized 0.1 Normal sodium thiosulfate solution until the yellow color has almost disappeared.

13. Add about 2 mℓ of starch solution and continue titration until the blue indicator color just disappears.

14. Record the volume of sodium thiosulfate solution used.

15. Calculate the iodine number as follows:

$$\frac{X}{M} = \frac{A - (2.2 \times B \times \text{m}\ell \text{ thiosulfate solution})}{\text{fixed carbon (g)}}$$

$$C = \frac{N_2 \times \text{m}\ell \text{ thiosulfate solution}}{50}$$

$$\text{Iodine Number} = \frac{X}{M} D$$

Where:

X/M = mg iodine adsorbed per g of carbon

$A = N_1 \times 12693.0$

$B = N_2 \times 126.93$

C = residual filtrate normality

N_1 = normality of iodine solution

N_2 = normality of sodium thiosulfate solution

D = correction factor (obtained from a residual filtrate Normalization graph using term C)

The capacity of a carbon for adsorbing any solute is dependent on the concentration of the solute in the liqid contacting the carbon. Thus, the concentration of iodine in the filtrate must be specified. The amount of carbon sample to be used in the determination is governed by the activity of the carbon. If the residual filtrate normality, C, is not within the range of 0.008N to 0.035N, the above procedure must be repeated using a different size sample. To maximize the accuracy of the test, the potassium iodide to iodine weight ratio should be 1.5 to 1.0 in the standard iodine solution.

Molasses Number

Using 100 as a standard for virgin Aqua Nuchar A, a comparative molasses number was determined according to the following procedure:

1. Prepare a stock molasses solution as follows:

   a. Dilute 20 m of molasses to 250 mℓ with distilled water.

   b. Filter the entire solution with Whatman GF/B 5.5 cm filter paper.

   (This is good for 24 hours only and should be refrigerated).

2. Weigh 0.50 g each of fixed carbon to be tested and virgin carbon (molasses number 100-arbitrary) previously dried at 105°C.

3. Add the samples to separate 150 mℓ beakers and pipette 50.0 mℓ of stock molasses solution to the beaker.

4. Stir and heat the mixture to 90°C and then cool.

5. Filter the samples through 5.5 cm Whatman GF/B filter paper.

6. Determine the absorbance of each filtrate with a spectrophotometer at 425 nm, using distilled water as zero absorbance.

7. Calculation:

$$\text{Molasses Number} = \frac{100 \times A_{100}}{A_{\text{Sample}}}$$

Where: A = absorbance

LABORATORY PROCEDURES

Equilibrium Adsorption Isotherm Test

The adsorptive response of soluble wastewater organics on virgin or regenerated carbon was determined using twenty-four hour grab composite samples of chemical treatment unit effluent. Contacting of the carbon was accomplished with a six place Turbitrol Jar Test Apparatus. All adsorption tests were conducted at a room temperature of 68 to 75°F. The step by step procedure followed was:

1. Acid rinse all glassware.

2. Prepare carbon samples as follows:

   a. Tare 100 mℓ beakers

   b. Add a desired weight of oven dried (105°C) fixed carbon to each of the beakers and then add 20 to 30 mℓ of distilled water.

   c. Cover beakers with watch glasses and place on a hot plate. Bring contents to a boil, remove from the hot plate and cool to near ambient temperature.

3. Add 1.0 to 1.5 liter samples of wastewater to each of six beakers and mix at 100 rpm.

4. Measure the initial temperature and pH of each sample, adjusting the pH to 7.0 to 7,5 if necessary with concentrated H_2SO_4 or NaOH.

5. To the mixing samples, affect a quantitative transfer of the degassed carbon (from Step 2), recording the amount of distilled water used for rinsing. A blank wastewater sample is always run without any carbon.

6. Mix samples for one hour at 100 rpm (preliminary tests had indicated that at least 95 percent SCOD removal was obtained at about 20 minute contact time).

7. At the end of one hour measure pH and temperature of each sample.

8. Filter a sample from each beaker through a 0.45 micron membrane filter and determine the COD concentration.

Standard procedure involved running triplicate COD tests on the blank and duplicate tests on each carbon contact sample. The difference between the blank and carbon contact average organic concentration was considered adsorbed organics. The organic removal (mg/ℓ adsorbed organics per mg/ℓ carbon) was plotted versus the average equilibrium organic concentration (mg/ℓ on log-log graph paper).

Gravity Thickening Tests

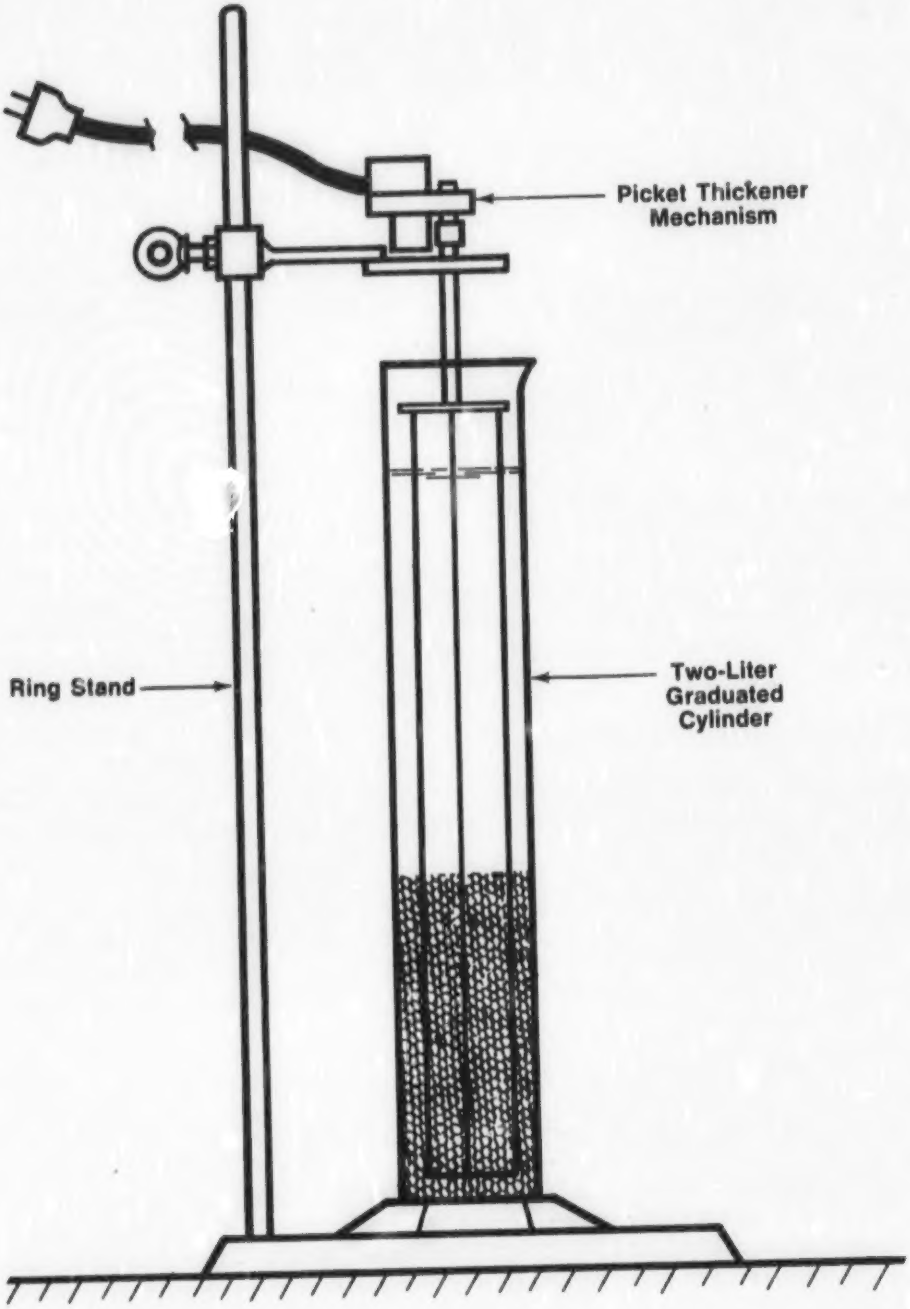
Laboratory gravity thickening tests were conducted by the following procedure. The laboratory test device used is shown on Figure A-1.

1. Obtain a 24 hour grab composite sample of sludge blowdown. If tests are to be run at different initial solids concentrations, then a sample of overflow was also obtained for dilution of the sludge.

2. Pour 2 liters of representative sludge into a 2 liter graduated cylinder for which the volume versus depth relationship is known.

3. Gently mix the contents to insure uniform suspension. A rubber stopper on the end of a rod works well. If a conditioning chemical is used, it is added during this gentle mixing.

FIGURE A-1: SLUDGE THICKENING TEST APPARATUS


Picket Thickener Mechanism
Ring Stand
Two-Liter Graduated Cylinder

4. Immediately insert the picket thickener mechanism into the sludge and start it rotating at 6 revolutions/hr. Also, immediately start a clock timer.
5. Observe and record sludge interface height (mℓ) at appropriate time intervals such that a smooth curve is identified. The test is normally run until no further thickening is observed for a several hour period.
6. Remove the picket mechanism and decant as much clear supernatant as possible, recording the remaining volume of sludge.
7. Filter all of the thickened sludge through a Whatman #1 filter paper, using a Buchner funnel. The filtrate is saved for a specific gravity determination.
8. Dry the filtered sludge at 105°C to constant weight to determine the mass of sludge solids (supernatant from Step 6 is assumed to contain no suspended solids).

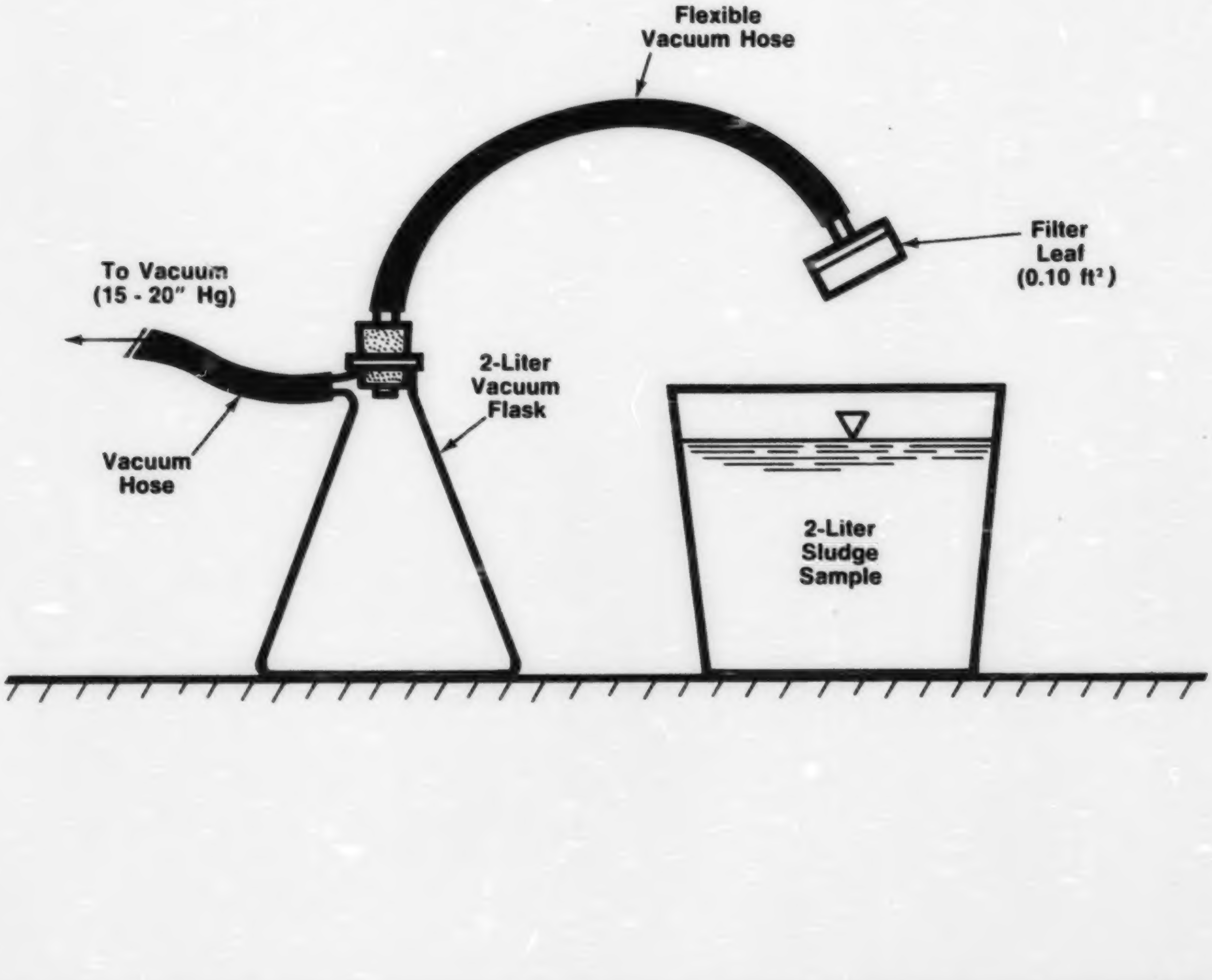
From the above data, slurry and solids densities can be calculated.

Vacuum Filter Leaf Tests

Leaf tests were conducted on composite sludge samples. The filter circular leaf used had an area of 0.10 sq ft. The general arrangement of laboratory equipment for conducting the leaf test is shown on Figure A-2. The procedure followed was:

1. Place about 1 to 2 liters of a representative sample of sludge in a 2 liter beaker.
2. If conditioning chemicals are used, they are gently mixed into the sludge with a large spatula.
3. Apply a vacuum to the filter leaf.
4. Immerse the properly sealed and conditioned filter leaf into the mixed sludge. The vacuum level is immediately adjusted to the desired level and left there for the desired cake form time (FT).

FIGURE A-2: SLUDGE DEWATERING TEST APPARATUS


Flexible Vacuum Hose
Filter Leaf (0.10 ft²)
To Vacuum (15 - 20" Hg)
2-Liter Vacuum Flask
Vacuum Hose
2-Liter Sludge Sample

5. Remove the leaf from the sludge and maintain the desired vacuum for a desired cake dry time (θ_D).

   a. If the cake cracks, note the time.

   b. Note the time of any reduction in vacuum level during dry time.

6. Turn off vacuum at end of dry time.

7. Note any difficulty in removing the cake from leaf.

8. Place all cake on a tared evaporation dish and weigh immediately; this will give wet filter cake weight (W_w).

9. Measure filter cake thickness (T_c).

10. Dry the filter cake to constant weight at 105°C; this will give the dry filter cake weight (W_D).

11. Determine the filtrate volume (contents of vacuum flask) and filtrate suspended solids and/or turbidity.

12. Determine sludge suspended solids of samples collected before and after addition of any conditioning chemicals such as lime.

The following are typical vacuum filter performance parameters which can be computed from the above data:

1. % Solids Capture $= 100 \times \frac{(\text{feed SS - filtrate SS})}{\text{feed SS}}$

2. % Moisture Content $= M_c = \frac{W_w - W_D}{W_w} = \frac{\text{wet weight-dry weight}}{\text{wet weight}}$

   (Note: dry weight includes any conditioning chemicals)

3. Filter Cake Weight $= W = \frac{\text{dry cake weight } (W_D)}{\text{leaf area}}$, in lb/sq·ft

4. Form Filtration Rate (FFR) $= \frac{\text{dry filter cake weight (W)}}{\text{form time (FT)}}$

   (report as Lb of dry solids/hr·sq·ft)

5. Full-Scale Filtration Rate (Y) = Yield

$$= \text{FFT} \times \text{Submergence}$$

$$= \frac{\text{FFR} \times \text{FT}}{\text{Cycle Time (CT)}}$$

$$= \frac{W}{CT}$$

Note: If conditioning chemical adds suspended solids (SS) to feed slurry, the FFR should be multiplied by the ratio of original feed SS to original feed SS plus added SS to obtain FFR and Y in terms of original feed SS.

6. Dry Cake Density $= \frac{\text{dry filter cake weight}}{\text{leaf area and cake thickness}}$

$$= \frac{W}{\text{cake thickness } (T_c)}$$

Typical vacuum filter performance parameter relationships for a given feed sludge are:

1. At constant vacuum:

$$FFR = K_1 \ (FT)^{-k}$$

Where: K_1 = a constant

k = a constant

2. At constant cake thickness:

$$FFR = K_2 \ (FT^{-z}$$

Where: K_2 = a constant

z = a constant (theoretically - 1.0)

3. At constant form time (FT):

$$FFR = K_3 \ (\Delta P)^{Y}$$

Where: K_3 = a constant

Y = a constant (theoretically for a noncompressible cake = 1/2)

APPENDIX B

REDUCTION OF LABORATORY GRAVITY THICKENING TEST DATA

All laboratory batch settling test data were reduced using a standard graphical technique employed by the Sanitary Engineering Technology Department, Eimco BSP Division of Envirotech Corporation, Salt Lake City, Utah. The basis for this technique is found in Kynch's theory.[17] This graphical technique is depicted in Figure B-1 and is made by the following procedure:

1. Draw a linear initial sludge interface subsidence slope as per line AB.
2. Draw a final linear slope after the knee of the curve as per line CD.
3. Construct a bisect (EF) of the obtuse angle at E, the intersection of tangent lines AB and CD.
4. Where the bisect (EF) crosses the sludge subsidence curve, at point G, construct line GH perpendicular to the bisect line EF and tangent to the curve.

Line GH is called the operating line and presumably identifies the interrelationship between predicted underflow concentration and thickener solids loading. Underflow suspended solids (g/ℓ) equals the grams of dry sludge solids present in the sample divided by the sludge interface height (liters). Thickener solids loading (TSL) at a desired underflow suspended solids concentration is computed as follows:

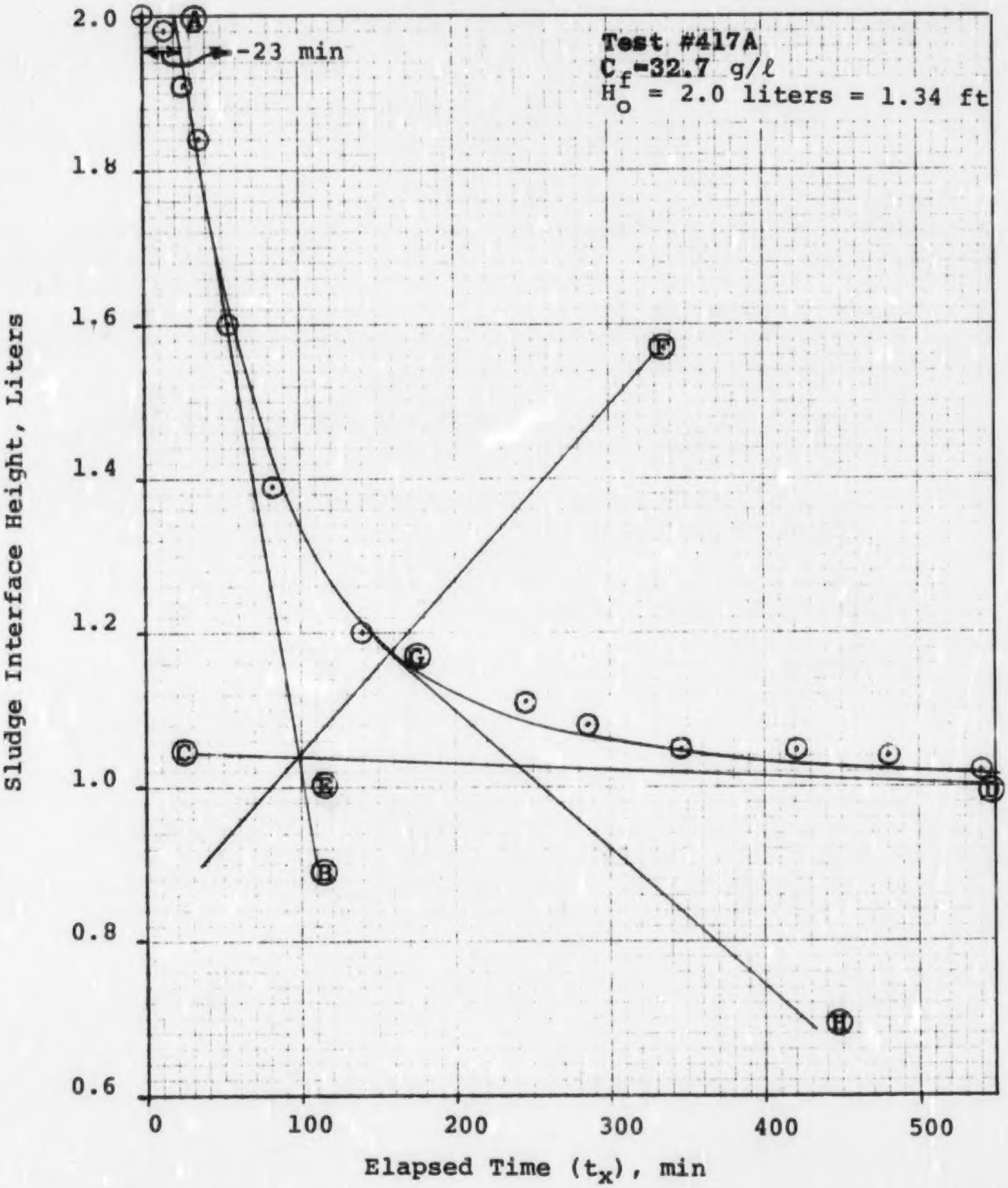
$$TSL = \frac{C_f H_o}{t_x}$$

Where:

C_f = feed or initial suspended solids in lb/cu ft

H_o = initial sample height in ft

FIGURE B-1:
GRAPHICAL TECHNIQUE FOR
LABORATORY GRAVITY THICKENING RESULTS



t_x = time in days related to the desired underflow suspended solids (interface height) by the operating line GH.

If, as on Figure B-1, there is a discontinuity in the subsidence curve near the starting interface height, a correction must be made to the observed t_x. Such a discontinuity is sometimes due to nonquiescent slurry conditions at elapsed time zero. For the data on Figure B-1, the correction is seen to be 23 minutes.

Laboratory thickening data reported in Section VII were obtained by making "graphical" constructions and computing the TSL for several underflow suspended solids concentrations and t_x's defined by the operating line GH. At t_x values less than point G, the actual subsidence curve was assumed to be the operating line.

Judgement is normally applied in choosing the plotting scale on Figure B-1. Normal procedure, when using a two liter graduated cycliner, is to plot the subsidence curve on 10 by 7 inch arithmetic graph paper. The ordinate, interface height, is the 10 inch scale and the abscissa, time, is the 7 inch scale. A critical judgement is the location of the actual curve on the graph (i.e., the choice of a time scale). The standard procedure used was to locate the curve such that the point G, defined as critical point, is located at about 33 percent of the 7 inch abscissa or at 2.3 inches. The effect of the relative location of the "critical point" on the abscissa on predicted TSL can be indicated by the following example.

On Figure B-1 the time scale is 100 min/in. resulting in the "critical point" being located at 1.6 inches on the 7 inch abscissa. The relationship between predicted TSL and desired underflow suspended solids, as defined by the operating line GH, is shown in Figure B-2 as curve B. This data on Figure B-1 was replotted twice, once at one-half and once at double the time scale shown. "Graphical" construction analysis for these two replottings resulted in predicted TSL versus underflow suspended solids relationship shown on Figure B-2 as curves A and C. Table B-1 shows pertinent parameters of this analysis for the three different plots. From Figure B-2, it is apparent that the choice of abscissa plotting scale has a pronounced effect on the predicted TSL at desired underflow suspended solids. The variation is greater as desired underflow suspended solids approaches the maximum obtainable concentration. Table B-2 indicates the substantial percent change in predicted TSL at various underflow concentrations.

FIGURE B-2:
REDUCED LAB THICKENING TEST DATA

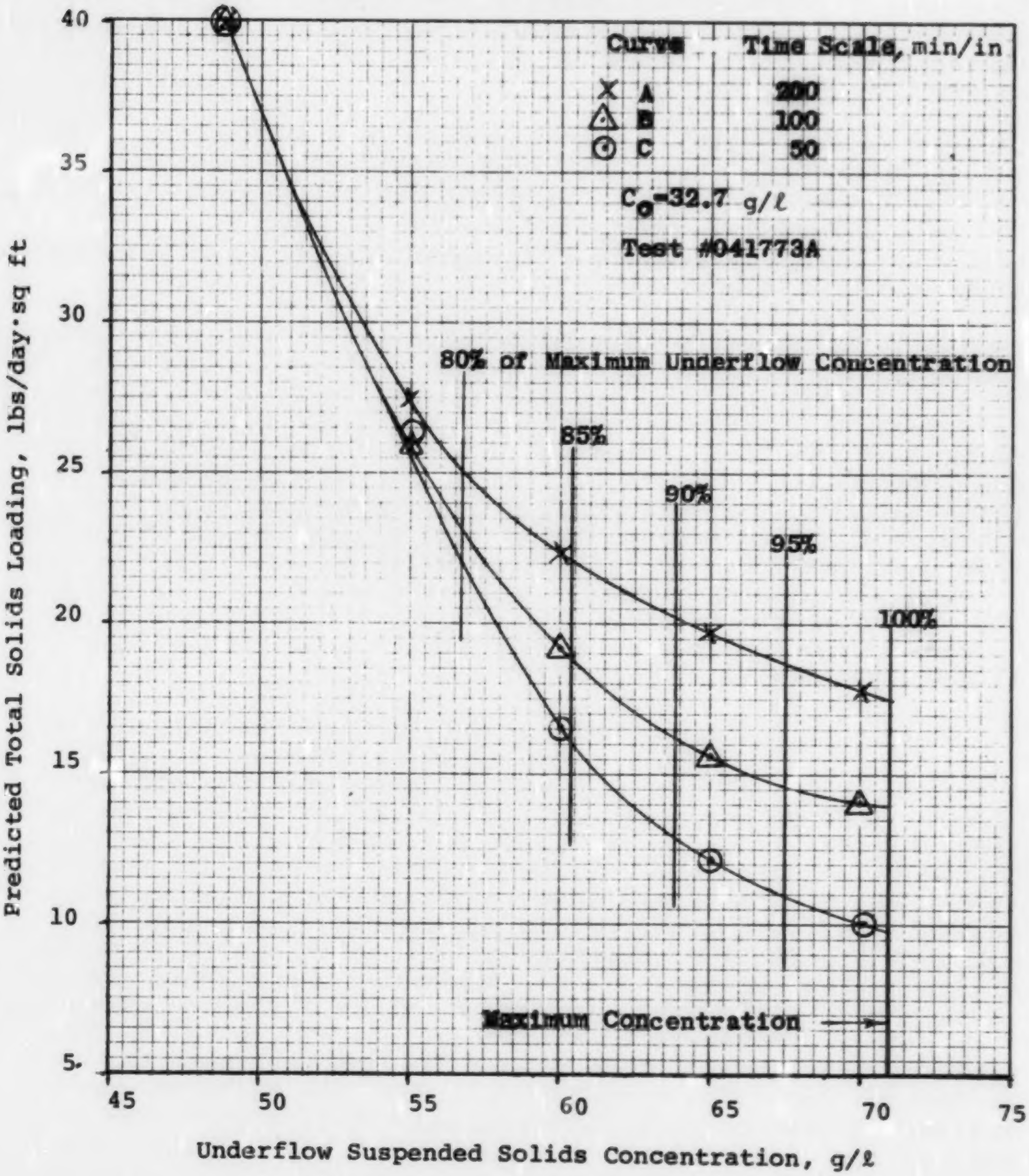


TABLE B-1

LABORATORY GRAVITY THICKENING DATA REDUCTION: GRAPHICAL CONSTRUCTION RELATIONSHIPS

	Curve Symbol (See Figure C-2)		
	A	B	C
Abscissa Scale, min/inch	200	100	50
Initial Slope:			
ℓ/min	-0.012	-0.012	-0.012
inch/inch	11.4	-6.20	3.13
T_x Correction, min	-23	-23	-23
Final Slope:			
ℓ/min	$-8x10^{-5}$	$-8x10^{-5}$	$-8x10^{-5}$
inch/inch	-0.080	-0.040	-0.020
Critical Time:			
min	200	162	137
inches	1.0	1.6	2.7
% of 7 inches	14	23	39
Operating Line Slope:			
ℓ/min	-0.0039	-0.0018	-0.00080
inch/inch	-0.99	-0.89	-0.80

TABLE B-2

ERROR IN PREDICTED THICKENER SOLIDS LOADING AT VARIOUS DESIRED UNDERFLOW CONCENTRATIONS

Underflow SS Concentration % of Maximum[a]	% Change in TSL per inch Deviation of Critical Time from 2.3 inches
80	6.1
85	13
90	17
95	19
100	17

[a] Maximum = 72 g/ℓ

An obvious conclusion based on the above example is that, if a conservative estimate of predicted TSL at a desired underflow suspended solids concentration is desired, then the abscissa plotting scale should be compressed (i.e., high values of min/inch). For this report, however, it is indeed unfortunate that a more rational and expedient method of laboratory thickening test data reduction was not available. It was essentially impossible to reduce over one hundred sets of thickening test data by hand and have all the critical points fall at a given location on the abscissa. Thus the laboratory results reported in Section VII have a data analysis error included.

APPENDIX C

UV LAB STUDY

A laboratory study was designed to identify some of the variables responsible for the poor correlation observed between UV absorbance and organic concentration. The specific variables considered were changes in the nature of the soluble organic fraction due to treatment selectivity effects and natural diurnal changes in relative organic concentrations.

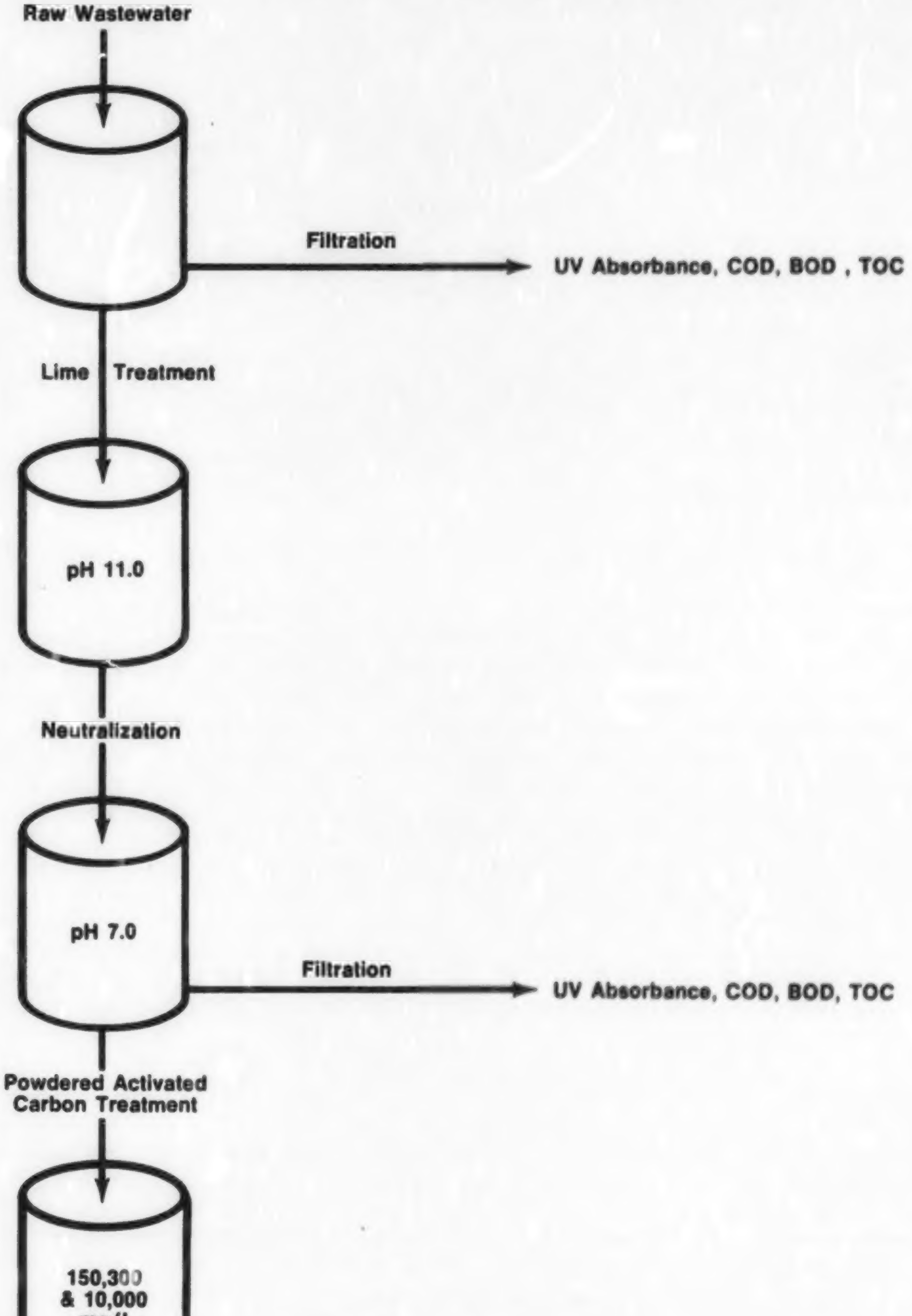
LABORATORY STUDY

Figure C-1 shows the approach used to simulate the PAC-PCT treatment effects. Laboratory simulation was necessary because this study was made during a period of pilot plant shut down when the project chemist had time available. Grab samples of raw wastewater were taken at high, average and low flow periods of the day. Organic concentrations of the wastewater were observed to be variable, low in the morning and high in the afternoon and evening. Each sample was treated with hydrated lime to a pH of 11.0 and settled. Supernatant was neutralized with H_2SO_4 to pH 7.0 and contacted with Aqua Nuchar A carbon at dosages of 150, 300 and 10,000 mg/ℓ. A sample from each treatment step (raw wastewater, lime treated, 150, 300 and 10,000 mg/ℓ carbon) was filtered through glass-fiber filter paper (GF/B) to remove particulate matter. The UV absorbance was determined for each sample at a wavelength of 254 nm. COD was considered the primary organic measure used because of testing accuracy and precision.

RESULTS

A plot of UV absorbance versus COD, BOD and TOC for each sample (high, medium and low organic concentration) is shown on Figures C-2, C-3, and C-4. From these plots, it is evident that UV absorbance versus organic concentration is not directly proportional for the entire process. An explanation of these nonlinear results may be possible by considering the four general types of organic material in

FIGURE C-1: LABORATORY STUDY


Raw Wastewater
Filtration
UV Absorbance, COD, BOD , TOC
Lime Treatment
pH 11.0
Neutralization
pH 7.0
Filtration
UV Absorbance, COD, BOD, TOC
Powdered Activated Carbon Treatment
150,300 & 10,000 mg/l
Filtration
UV Absorbance, COD, BOD, TOC

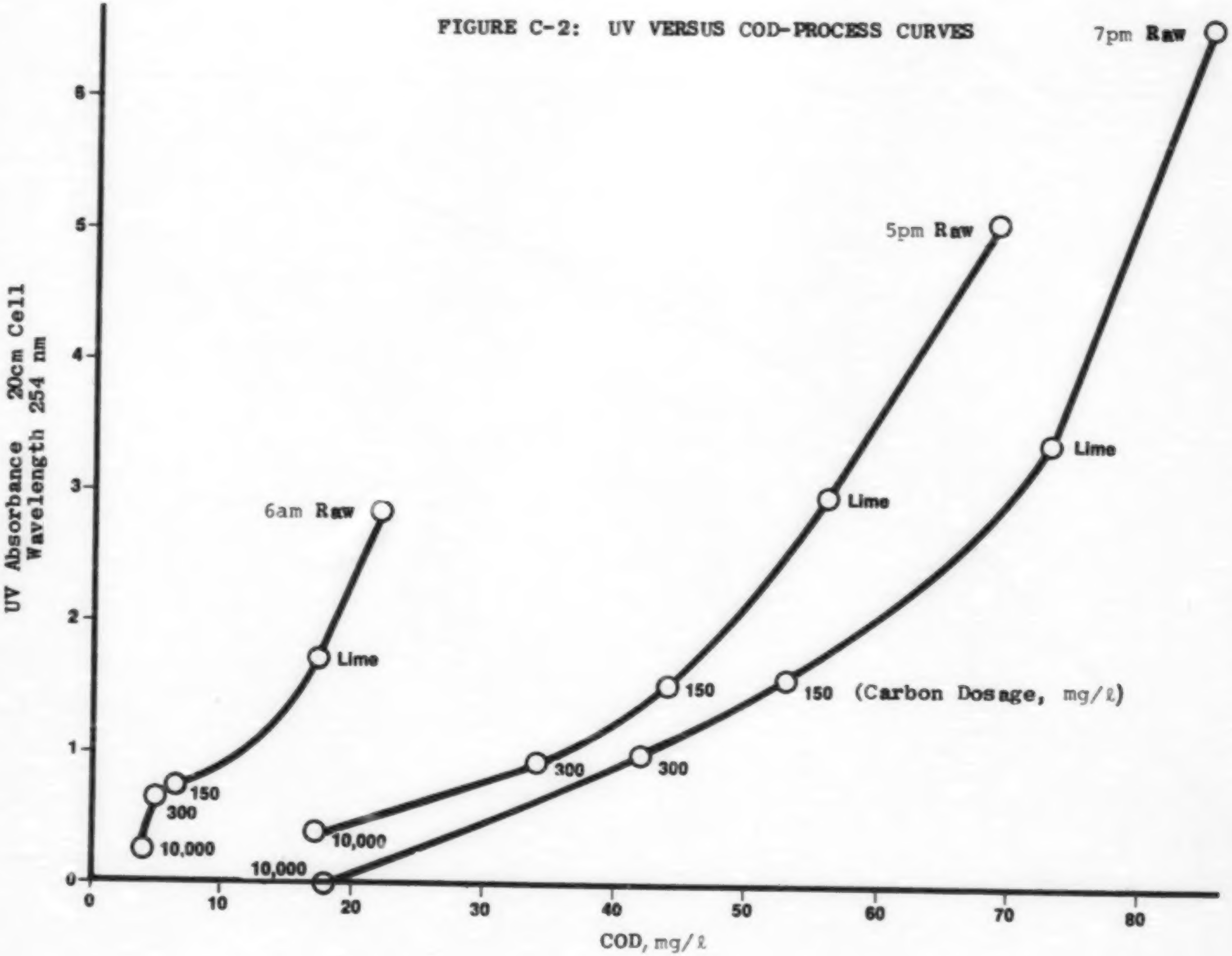


FIGURE C-2: UV VERSUS COD-PROCESS CURVES

FIGURE C-3: UV VERSUS BOD PROCESS CURVES

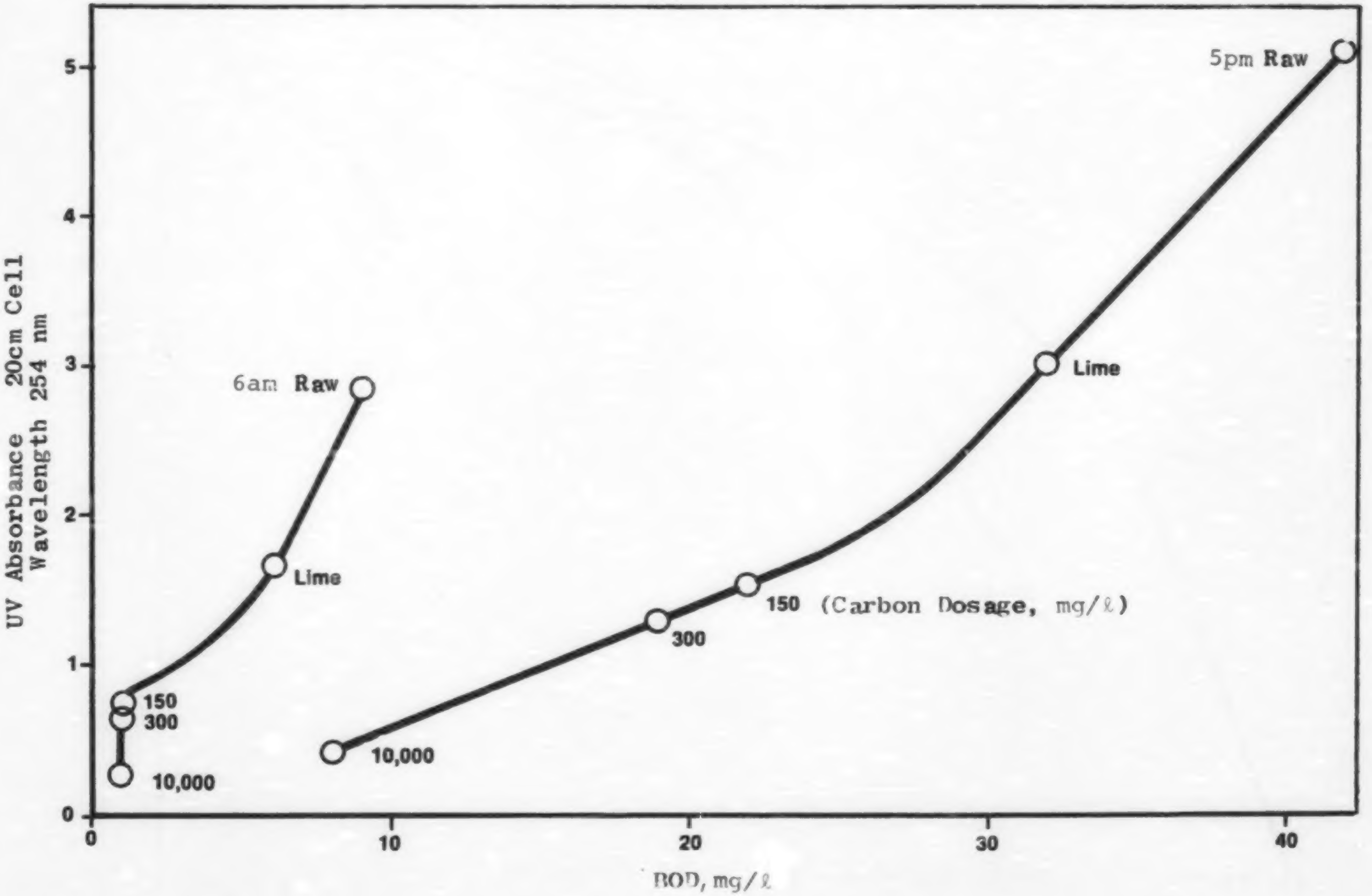
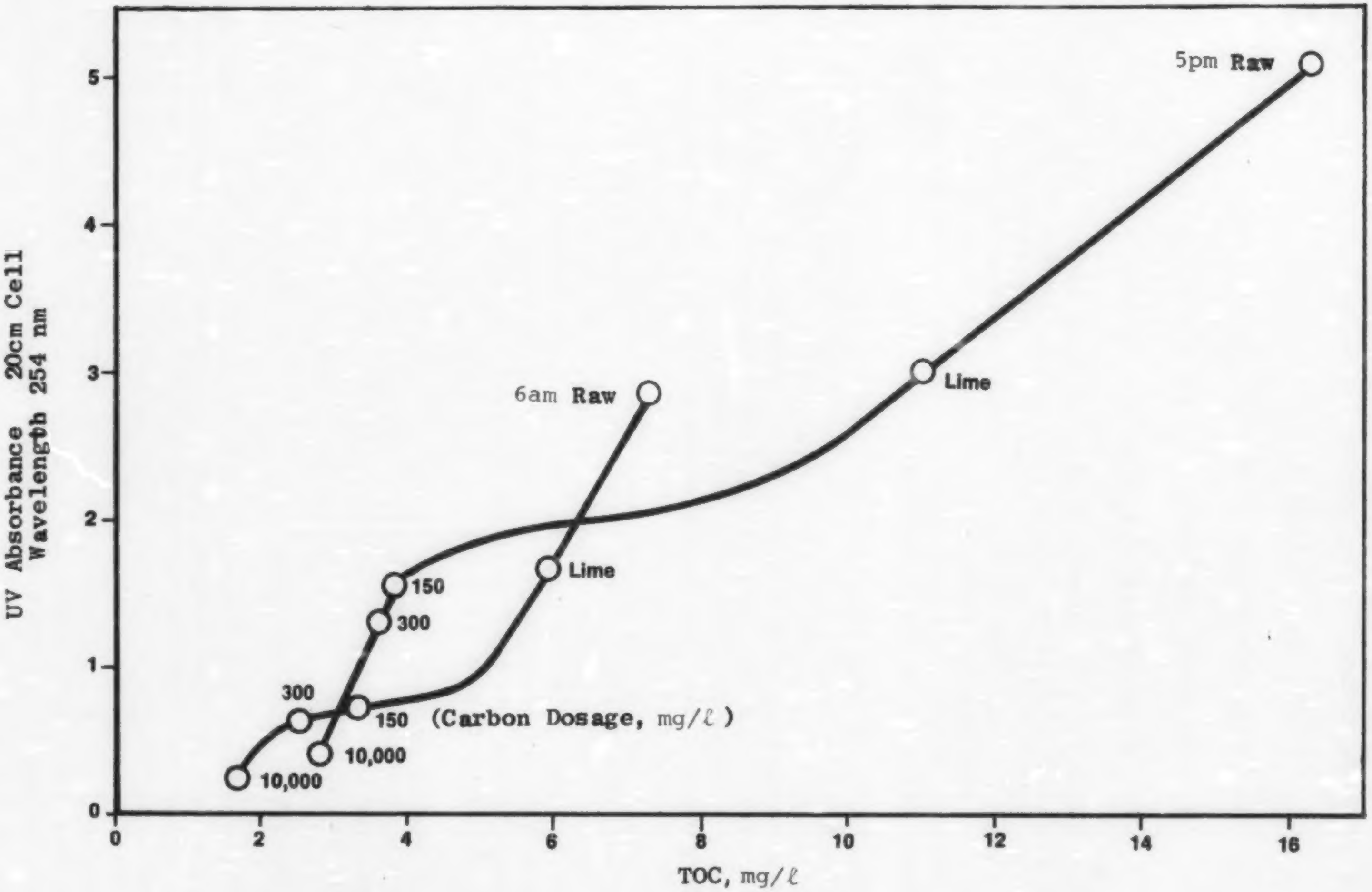


FIGURE C-4: UV VERSUS TOC PROCESS CURVES



the fraction of the wastewater which will pass through a GF/B glass-fiber filter paper:

1. Colloidal solids
2. Solutes easily adsorbed on carbon
3. Solutes difficult to adsorb on carbon
4. Solutes unadsorbable on carbon

The change in UV adsorbance with the change in COD from raw wastewater to lime treated water is probably due to selective removal of organic colloids. The effect of particulate organics has been shown by using various filter pore sizes and comparing UV adsorbance, TOC and turbidity.[14] In the present study, coagulation-flocculation probably removed some soluble organic material and thus the UV absorbance.

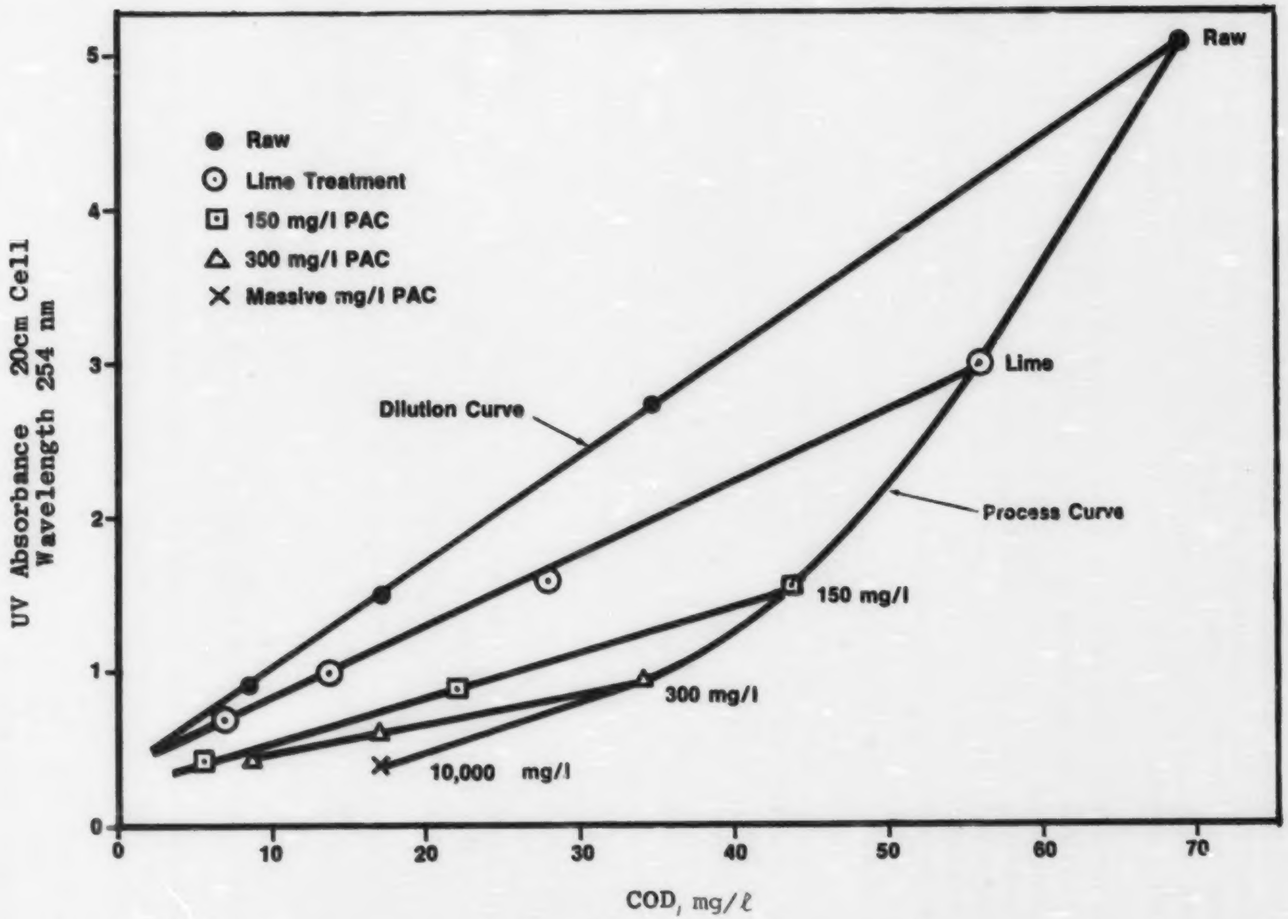
The knee of the curves, on Figures C-2, C-3 and C-4 is located in a transitional region between colloid removal and difficult to adsorb solute removal. The transitional region was caused by the preferential removal of easily adsorbed solutes. As the organic solutes become increasingly difficult to adsorb, the slope of the process curves continue to decrease until the slope becomes fairly constant approaching the unadsorbable solute region. The unadsorbable solutes, which vary in concentration with time, are probably responsible for a considerable variation in UV adsorbance with respect to organic concentration. This variation is evidenced by the massive carbon dosage data shown on Figure C-2. One sample containing 18 mg/ℓ COD showed a UV absorbance of zero while another sample containing only 4 mg/ℓ COD showed a UV absorbance of 0.25

It would appear that the nature of the soluble organic fraction changes significantly due to treatment effects, thus, making it virtually impossible to define a precise general correlation.

TREATMENT SELECTIVITY

Evidence of the changing nature of the soluble organic fraction and of the selectivity of carbon and coagulation-flocculation was demonstrated with the use of a consecutive dilution technique. Consecutive dilutions were made of each sample (raw wastewater, lime treated, 150, 300, 10,000 mg/ℓ carbon) in a process curve and a UV absorbance was determined for each dilution. A plot of UV absorbance versus COD on Figure C-5 shows a linear response for each sample. By

FIGURE C-5: SELECTIVITY DUE TO TREATMENT EFFECTS



reducing the COD concentration and not varying the relative concentration of the different organic solutes, the UV absorbance varied directly with respect to COD. By reducing COD concentrations by coagulation-flocculation and carbon contacting, a nonlinear curve (Figure C-2) resulted, indicating the selective effect of coagulation-flocculation and carbon treatment. This selectivity is responsible for the inability to achieve a correlation between UV absorbance and organic concentration for the entire process.

Figure C-6 indicates a UV-COD concentration correlation is possible by considering the specific treatment points (i.e., lime treatment, 150 mg/ℓ or 300 mg/ℓ carbon) of Figure C-2. However, it must be kept in mind that this is only a working correlation for this particular wastewater after specific treatment effects rather than a true general UV-organic concentration correlation.

NATURAL RELATIVE ORGANIC CONCENTRATION CHANGES

While UV scanning various raw wastewater samples, it was noted that the knee of the curve and the plateau region shifted (Figure C-7). The curve shift was probably due to natural diurnal changes in relative concentration of organic solutes. To confirm this hypothesis, a test was designed to demonstrate curve shift. Various concentrations of detergents (ABS type) and urine were scanned. ABS and urine were used because they can both be present in municipal wastewater. Typical results, as presented in Figure C-8, indicate detergents show a knee in the curve from 235 nm to 250 nm. Urine shows a knee from 250 nm to 270 nm. Various detergent-urine concentration ratios were scanned. They showed that as the detergent concentration increased relative to the urine concentration, the curve shifted to the left putting the standard 254 nm wavelength well within the plateau region. But as the urine concentration increased relative to the detergent concentration, the curve shifted to the right putting the standard 254 nm wavelength on the knee of the curve (Figure C-9). By so doing, the probability of obtaining a precise UV absorbance-organic concentration correlation is greatly reduced. Although detergents and urine are not the only species adding to the UV absorbance character of the wastewater, the above tests do demonstrate the effect of changing relative concentrations of typical wastewater organics on UV absorption characteristics.

DISCUSSION

The basic problem of achieving a general UV adsorbance-organic concentration correlation results from the changing relative concentrations of the organic solutes present in

FIGURE C-6: SPECIFIC TREATMENT POINT CORRELATION

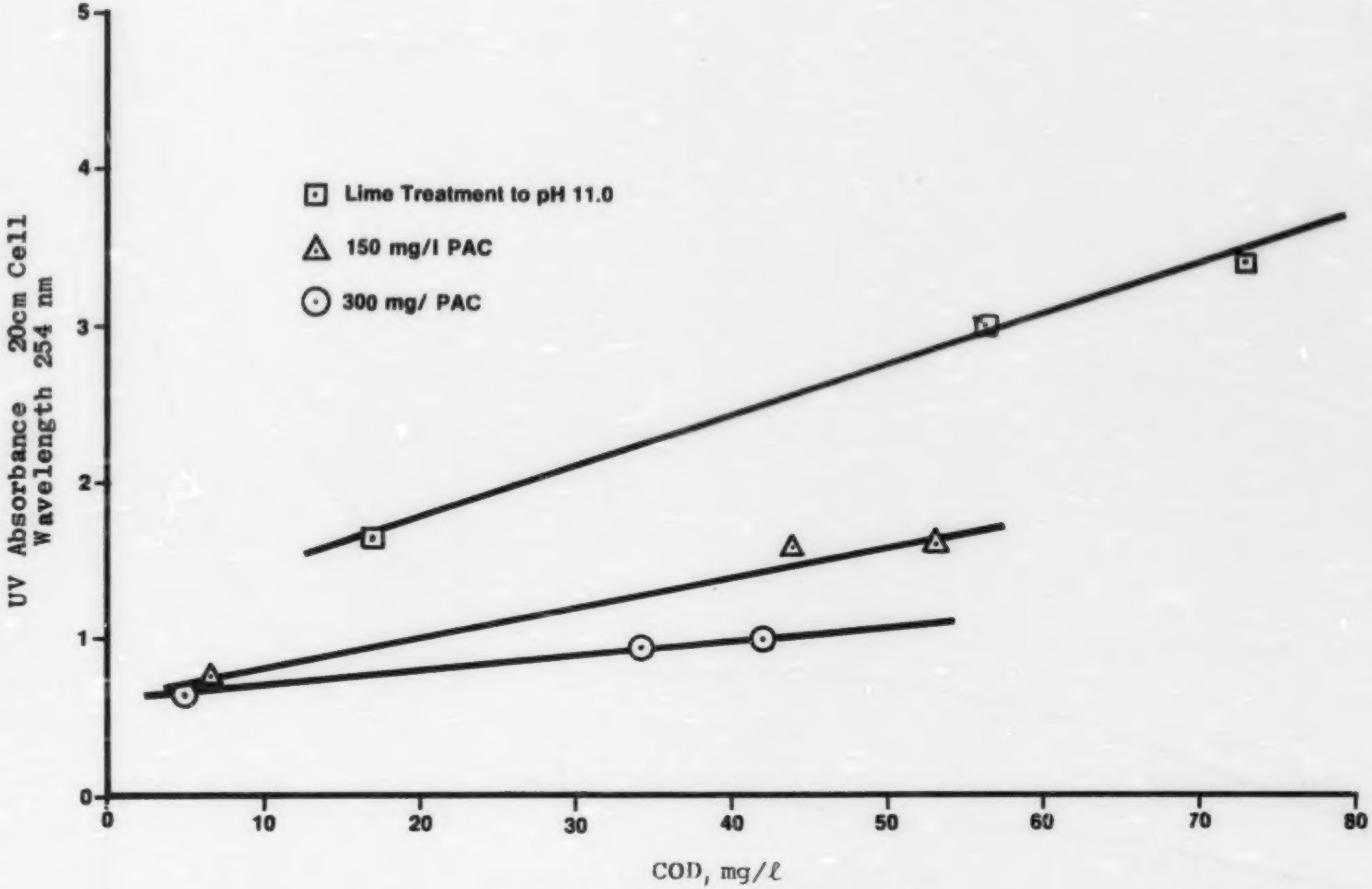


FIGURE C-7: DIURNAL CHANGES IN THE NATURE OF ORGANIC CONCENTRATIONS

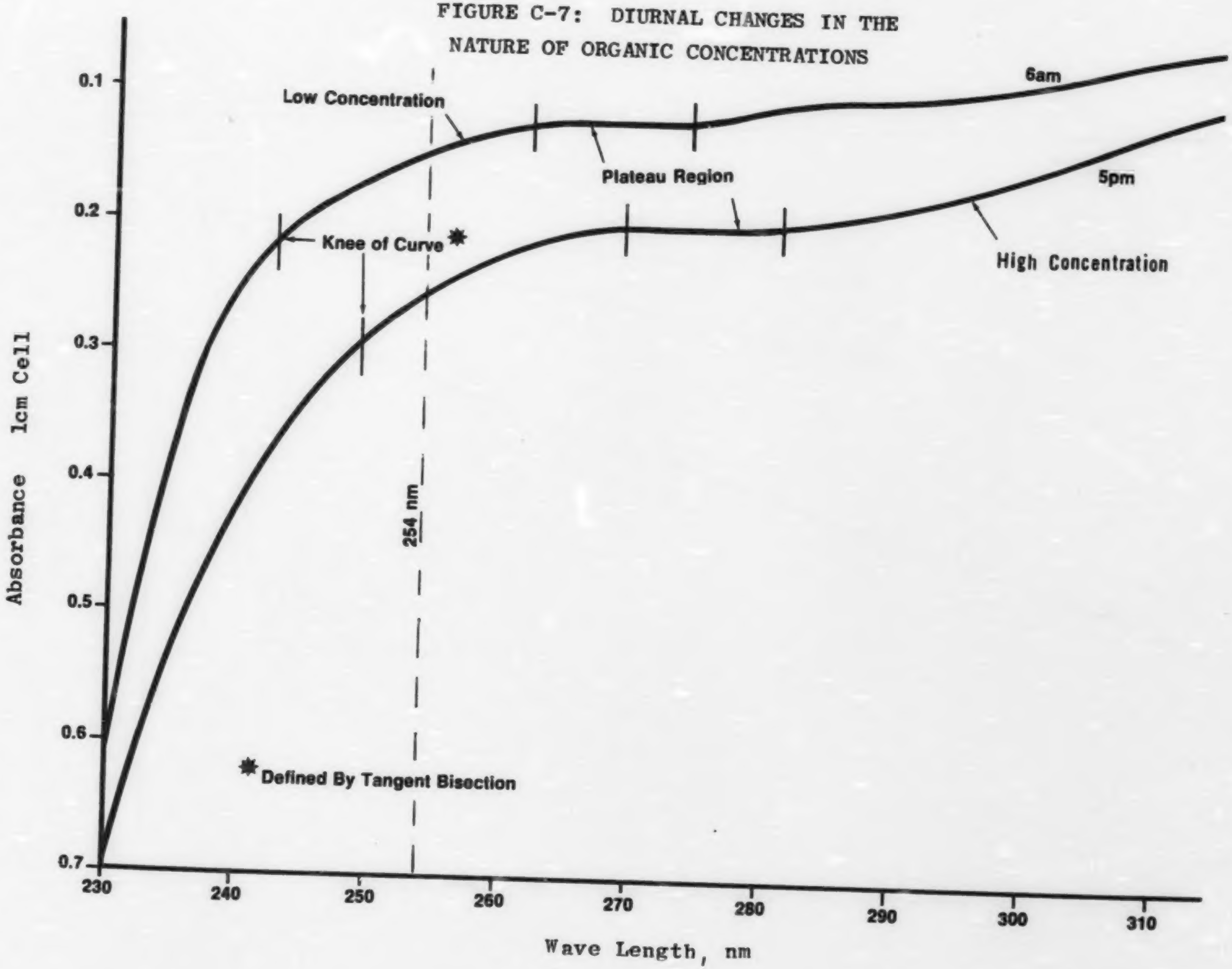


FIGURE C-8: ABS AND URINE UV SPECTRA

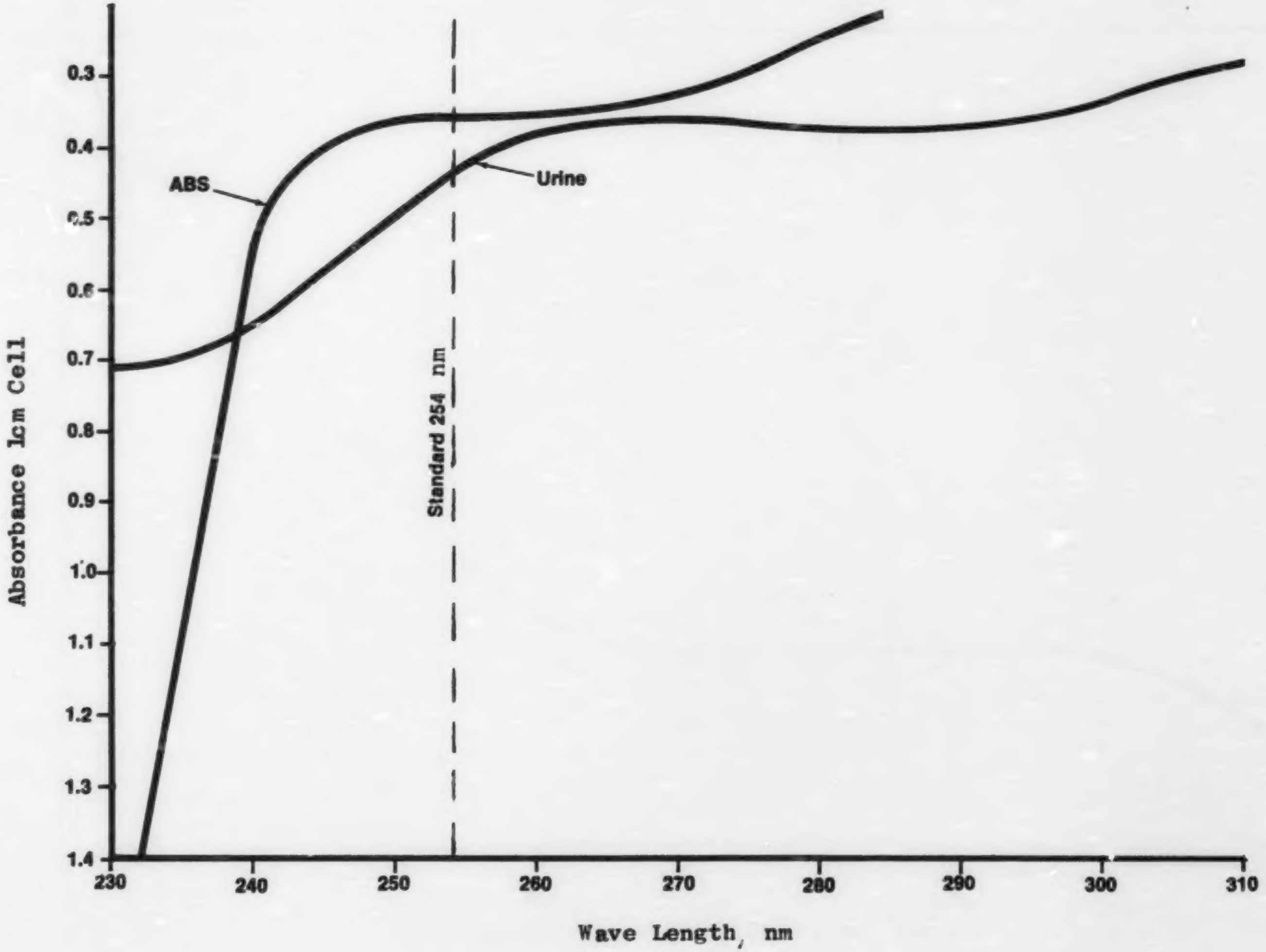
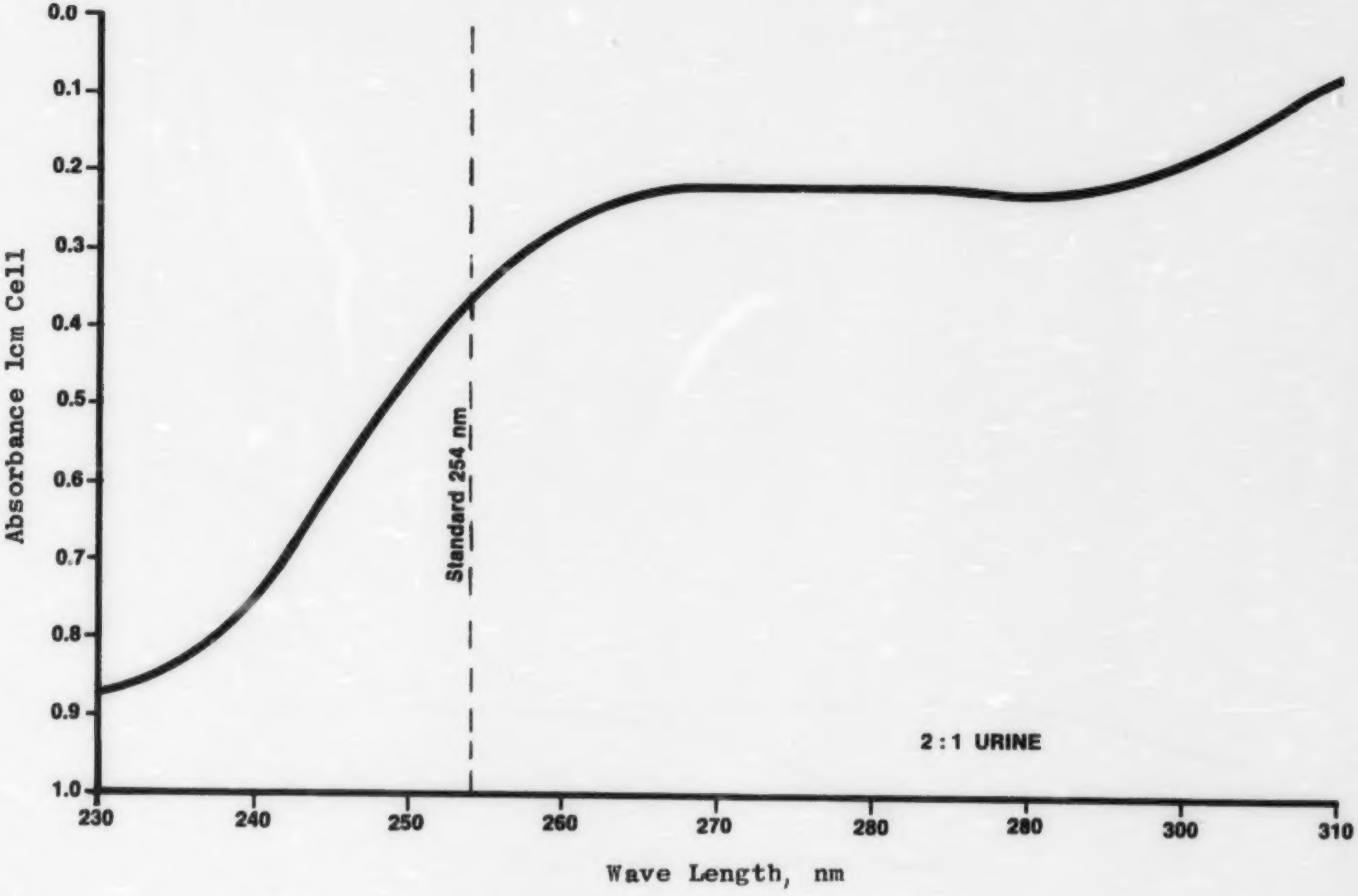


FIGURE C-9: UV-SPECTRA SHIFT DUE TO CHANGE IN RELATIVE ORGANIC CONCENTRATION



the wastewater. In the PAC-PCT process, relative concentration changes result from at least three basis causes:

1. Chemical treatment selectivity,
2. Carbon adsorption selectivity, and
3. Natural diurnal organic concentration changes.

Application of UV absorbance to monitor organic concentration is dependent on the treatment scheme used and the nature of the wastewater evaluated. If the wastewater or treatment process stream's relative organic concentrations are fairly constant and a nonselective treatment process (i.e., gravity clarification) is used, then a precise UV absorbance-organic concentration correlation might be possible for the process influent and effluent. In a selective treatment process, such as PAC-PCT, general UV absorbance-organic concentration correlation prospects are minimal. However, as previously shown a working correlation should be possible at specific treatment points (i.e., lime treated, 150, or 300 mg/ℓ carbon) if very consistent treatment effects are achieved.

As reported in Section IX an attempt to determine such a working correlation between COD and UV resulted in a less than satisfactory response.